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EVALUATION OF INNOVATIVE ALTERNATIVES FOR THE WEST WINDSOR PCP UPGRADE

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EXECUTIVE SUMMARY

Twenty-four Ontario municipal primary wastewater treatment plants bordering on the Great Lakes will be required to provide Best Available Treatment Economically Achievable (BATEA) in the near future. At the least, these plants will require secondary biological treatment. The City of Windsor's West Windsor Pollution Control Plant (PCP) is the largest of these facilities and it is expected that the upgrade will require a large capital and operating investment from the City. The West Windsor PCP's average hydraulic design capacity is 159,000 m^3/d and its peak hydraulic capacity is 350,000 m^3/d.

The existing West Windsor PCP uses chemical precipitation and flocculation to remove phosphorus and obtain an enhanced primary effluent. The City, expecting to be mandated to provide secondary treatment or BATEA, has investigated alternative treatment processes that would use the existing enhanced primary effluent to the greatest degree possible, while meeting the future plant's effluent requirements. Alternative treatment options would also have to be cost effective. The plant treated a yearly average wastewater flow of 116,000, 118,000, 139,000 and 120,000 m^3/d from 1988 to 1991. Based on overall data from the past 4⅓ year, the plant removed 82 and 64 percent of suspended solids (SS) and total BOD_5. The plant has met its monthly effluent total phosphorus (TP) requirement of 1.0 mg/L for all but one month over the same 4⅓ year period. The effluent had average concentrations of 24, 36 and 0.6 mg/L for SS, total BOD_5 and TP for the 4⅓ year period. Table E.1 summarizes the historical performance of the West Windsor PCP.

Table E.1
Performance Summary of West Windsor PCP

Period	Treated Flow (10^3 m^3/d)	Average Concentration (mg/L)					
		Influent			Effluent		
		TSS	$TBOD_5$	TP	TSS	$TBOD_5$	TP
Overall Period (January 1988 to April 1992 inclusive)	124	134	101	3.8	24	36	0.6
Yearly Averages							
1988	116	119	92	3.9	26	36	0.7
1989	118	124	95	4.0	24	39	0.7
1990	139	150	98	3.8	20	34	0.6
1991	120	138	113	3.7	26	38	0.6
Partial 1992 (to April 1992)	139	162	128	3.7	20	34	0.4

Notes: TSS: total SS
$TBOD_5$: total BOD_5

The plant obtains enhanced primary treatment through the addition of ferric chloride or alum for precipitation and coagulation, in conjunction with an anionic polymer for flocculation. According to Ministry of the Environment (MOE) typical values, the existing plant produces more sludge with a higher percentage of solids than a typical plant, using the same amount of chemical (or slightly less). The West Windsor PCP uses 12.1 mg/L of ferric chloride (as Fe^{+3}) or 4.6 mg/L of alum (as Al^{+3}) with polymer compared to MOE typical values of 16 and 9.1 mg/L for ferric chloride (as Fe^{+3}) or alum (as Al^{+3}). For the period January 1990 to June 1991, the West Windsor PCP produced 748 t/mo of dry solids.

The objectives of this study were as follows:

- To evaluate, at pilot scale, four alternative processes capable of providing secondary treatment to the West Windsor PCP
- To use the existing physical/chemical plant to the maximum degree possible
- To determine potential effluent requirements for the future plant
- To determine loading criteria for each process
- To gain operating experience with the four processes
- Communicate ongoing pilot plant performance results to Steering Committee

The overall study was conducted from March 1990 to April 1992.

Input from the MOE was sought to determine the potential effluent requirements for the upgraded facility. The MOE indicated to the City that, at the present time, the expected effluent design requirements for the upgraded plant would not be more stringent than the following:

- 15 mg/L SS
- 15 mg/L total BOD_5
- 0.5 mg/L TP
- 0.1 mg/L un-ionized ammonia-N.

The un-ionized ammonia-N (NH_3-N) effluent requirement would be based on the end-of-pipe discharge conditions in terms of wastewater temperature and pH. After project initiation, the Steering Committee recommended a target effluent total NH_3-N concentration of 2 mg/L be used to examine actual nitrification performance. During a follow-up extension period the total ammonia-N target was reassessed at 3 mg/L.

The four treatment processes that were operated at pilot scale were:

- Trickling Filter/Solids Contactor (TF/SC)
- Rotating Biological Contactor (RBC)
- Biological Aerated Filter (BAF)
- Modified Activated Sludge Process (ASP).

The modified ASP was designed to operate at a reduced aeration tank hydraulic retention time compared to Ministry of the Environment design guidelines. The four pilot plants were operated for about one year at various loads and influent wastewater temperatures. The TF/SC and the BAF processes were operated for an additional year to investigate outstanding issues.

Each pilot plant used a portion of the effluent from two of the full-scale primary clarifiers. The pilot plant influent concentrations averaged 23, 43, 0.9, and 9.4 mg/L for SS, total BOD_5, TP, and NH_3-N over the first year of the study period (July 2, 1990 to May 31, 1991). During part of this period, the influent concentrations were increased from October 30, 1990 to January 11, 1991 by discontinuing the polymer addition to the two full-scale primary clarifiers whose effluent was used as influent by the pilot plants. This change was made to increase the organic loading to the pilot plants, to assist in starting the solids contact portion of the TF/SC, and to determine the impact on the other processes. During this period, the influent SS and total BOD_5 increased to 31 and 46 mg/L, respectively. During the extended period of operation of the TF/SC and BAF units from September 1991 to April 1992 the pilot plant influent concentrations averaged 24, 47, 0.6, and 9.9 mg/L for SS, total BOD_5, TP, and NH_3-N.

During the pilot plant testing, each of the process loading and operational parameters were manipulated resulting in a period when each process was optimized. These periods are shown in Table E.2.

Table E.2
Optimization Periods for Operation of Each Pilot Process

Pilot Process	Periods of Optimization
TF/SC	• December 18, 1990 - May 1, 1991 • June 1, 1991 - August 2, 1991 • September 17, 1991 - April 26, 1992
RBC	• July 5, 1990 - May 31, 1991
BAF	• August 20, 1990 - August 2, 1991 (excluding period January 11, 1991 - March 11, 1991) • January 15, 1992 - April 26, 1992
ASP	• November 16, 1990 - May 31, 1991 (excluding the period December 22, 1990 to January 11, 1991)

Operation of the TF/SC was interrupted on May 1, 1991 because of a loss of the solids inventory from the SC unit. The SC unit was restarted in June 1991 by increasing the loading to the TF tower, in an attempt to increase the loading to the SC unit. Nitrification was interrupted in the BAF due to a series of vigorous backwashes that were conducted in the Fall of 1990 to investigate hydraulic limitations through the unit. The period when attempts were made to re-establish nitrification by reducing the hydraulic loading and seeding the unit with nitrifying bacteria from the TF/SC was excluded from the analysis of the BAF. The operation of the ASP was interrupted by cold weather operational problems. The TF/SC process was operated for an additional year to investigate the operation and viability of the SC. The BAF process was operated for an additional year to investigate its ability to nitrify and handle peak flow events.

The performance of the pilot plants over the study period is summarized in Table E.3 with respect to:

- Effluent performance on a monthly and daily basis compared to the target effluent objectives and an upper limit based on MOE Policy 08-01.
- Ability to handle peak daily and diurnal flows on a hydraulic basis.
- Solids handling

EFFLUENT PERFORMANCE

- the best performance in terms of meeting the effluent design objectives was obtained from the BAF and RBC units. Both of these processes experienced periods when monthly total effluent NH_3-N concentration was greater than 2.0 mg/L. The BAF's effluent was tertiary quality, with the effluent SS less than 5 mg/L for 71 percent of the time.

- BAF nitrification was interrupted when potential hydraulic limitations were investigated and vigorous backwashing of the unit was initiated. During the extended operation period, the BAF was able to nitrify consistently in the range of 2 to 3 mg/L effluent NH_3-N concentration and handle peak flows after the media depth was increased from 2.0 to 2.5 m.

- The RBC unit periodically exceeded the effluent total NH_3-N target of 2.0 mg/L, but easily met the required effluent objective concentrations. Intentional process changes to achieve 2 mg/L NH_3-N were not made during cold weather months. The RBC effluent would meet a 3 mg/L effluent ammonia-N concentration, at loadings used in this study for the majority of the time.

- The ASP obtained the lowest effluent total NH_3-N concentrations which, based on a monthly average, were always less than 2 mg/L. The ASP also achieved the design objective for BOD_5, SS, and TP for 83 percent of the time.

- The TF/SC process nitrified well, but did not consistently meet the effluent objectives for SS, total BOD_5, and TP on a monthly basis.

Table E.3
Overall Summary of Pilot Plant Performance

Comparison Item	Pilot Plant Process			
	TF/SC	RBC	BAF	ASP
Monthly Compliance (%): Based on the percentage of months units operated with effluents less than design objectives. Objectives were 15, 15, 0.5, 0.1 mg/L SS, total BOD_5, TP, and un-ionized ammonia-N. Also target ammonia-N concentrations of 2 and 3 mg/L for process operation.				
• SS	66	100	100	83
• Total BOD_5	50	100	100	83
• TP	83	100	100	83
• Un-ionized NH_3-N	100	100	100	100
• Total NH_3-N				
2 mg/L	100	40	40	100
3 mg/L	100	90	80	100
Daily Performance (%): Percentage of time effluent from pilot plants was less than design objectives. Also target ammonia-N concentrations of 2 and 3 mg/L for process operation.				
• SS	55	96	98	75
• Total BOD_5	51	81	85	68
• TP	64	92	94	81
• Un-ionized NH_3-N	100	100	100	100
• Total NH_3-N				
2 mg/L	82	55	53[1]	90
3 mg/L	92	74	71[1]	95
Daily Performance Upper Limits (%): Percentage of time effluent from pilot plants was less than upper limit concentration. In this case 25, 25, and 1.0 mg/L for SS, total BOD_5, and TP (MOE Policy 08-01).				
• SS	92	99	99	92
• Total BOD_5	89	99	98	95
• TP	98	98	>99	97
Hydraulic Performance (Y - no difficulty, U - unsure): Ability of pilot plants to handle peak day and diurnal flow variations.				
• Can handle diurnal flow variations	Y	Y	Y	Y
• Can handle peak day flows	Y	Y	Y	Y
Solids Handling (T - typical, P - problem): Characteristics of sludges generated by pilot plants.				
• Solids Yield	T	T	T	T
• Solids Settling	T	T	T	T
• Solids Dewatering	T	T	T	T

Notes:

[1] BAF's performance with respect to ammonia-N removal based on performance prior to vigourous backwashing (September 7, 1990 to October 31, 1990) and on the period January 22, 1992 to April 26, 1992 after media depth increased to 2.5 m.

[2] ASP pilot plant did not operate at elevated flows, but City's experience at the Little River PCP indicates process can handle 2.2 peak factor.

HYDRAULIC PERFORMANCE

- The TF/SC and the RBC units had no hydraulic difficulty handling peak day or diurnal flow variations. Peak day flows were based on 2.2 times the average daily flow, based on the rated peak factor of the existing full-scale plant.

- The ASP treated the diurnal flow pattern five days a week, but only handled a 1.5 times peaking factor. This peaking factor resulted because the unit was not operational during the majority of the wet weather peak day hydraulic testing conducted in the fall of 1990. Based on the City's experience at the Little River PCP activated sludge facility and the performance of the ASP, this low hydraulic peaking factor is not expected to be a full-scale plant limitation.

- The BAF initially appeared to have hydraulic limitations late in the fall of 1990. The investigation of these problems and intensive backwashing led to the loss of nitrification in the unit. There were indications that media was syphoned from the unit during high flows due to the pilot plant design. Since the pilot plant did not have automatic level recording and backwash capability, it was unknown whether the unit always treated high flows. However, during the extended operation period, the BAF was able to handle 2.2 times the average daily flow over a 3-day period. During this test the unit was manned 24-hours per day.

SOLIDS HANDLING

- All the pilot plants solids production varied from 0.6 to 0.8 kgVSS/kg total BOD_5 removed. All of the units were comparable in terms of solids handling, settling, thickening and dewaterability.

OVERALL CONCLUSIONS

To relate each of the processes, a comparative evaluation process was used. The approach evaluated each of the processes with respect to five categories. These categories were effluent performance, hydraulic performance, solids handling, operational aspects, and economics. The maximum contribution for each category was 36, 7, 9, 18, and 30 percent as a percentage of the total rating. Effluent performance, hydraulic performance, and solids handling were rated based on the results shown in Table E.3. Operational aspects were rated based on an evaluation completed by members of the Steering Committee.

The economic evaluation is based on comparative full-scale process cost estimates developed by LaFontaine, Cowie, Buratto and Associates Ltd. Overall, each of the four processes tested at West Windsor are expected to be more economical than a conventional activated sludge plant. Capital costs for each process varied from 76 to 89 percent of an equivalent conventional ASP. However, the capital and O&M costs for each process are expected to be the same at this level of accuracy.

Based on the comparative evaluation:

- The BAF process scored highest for the West Windsor PCP upgrade, followed in order by the RBC, the ASP, and the TF/SC. The BAF scored highest because of its excellent performance in terms of SS, total BOD_5 and TP, and due to the resolution of the process's ability to nitrify and handle peak flows. The RBC scored well because of its overall consistent performance with respect to all effluent objectives and its ease of operation. The ASP had the advantage of being a process that is familiar to the City. The evaluation results for the RBC and ASP units were marginally less than the BAF.

- The BAF obtained tertiary effluent quality. If the effluent ammonia-N requirement was 3 mg/L, the BAF pilot plant would be the optimal alternative. An effluent NH_3-N requirement of 3 mg/L would be sufficient to meet the effluent un-ionized NH_3-N requirement of 0.1 mg/L based on West Windsor wastewater characteristics.

- The TF/SC scored lowest primarily because of its inadequate effluent performance, especially in terms of SS, TP and total BOD_5, as well as the uncertainty associated with the operation of the SC unit. The SC unit was difficult to start up and maintain, and the unit experienced a solids washout in May 1991. Therefore, it is concluded that the coupled TF/SC process, based on its effluent performance during the pilot testing, not be considered as an alternative for the plant upgrade without additional testing, including conventional primary treatment prior to the TF/SC, or operating the unit in a different configuration such as an activated biofilter process.

- With confidence, the RBC, BAF, and ASP can provide the required effluent objectives on a consistent basis.

REPORT RECOMMENDATIONS

Overall, the report recommends the following:

- that to provide consistent nitrification, a 2.5 m media depth be used for the BAF process
- that the loading rates from this pilot study be used as a basis for designing any of the four treatment processes evaluated
- that the coupled TF/SC process, based on its effluent performance during the pilot testing, not be considered as an alternative for the plant upgrade without additional testing, including conventional primary treatment prior to the TF/SC or operating the unit in a different configuration such as an activated biofilter process
- that, if the City decides to proceed with either the TF/SC (or other TF configurations), BAF or RBC processes, further full-scale site inspections and plant data be reviewed to assist in familiarizing the City with these processes
- that, to obtain a complete understanding of the performance and viability of the BAF process, the City inspect a number of full-scale BAF facilities which have been operating for several years. This is essential to gain a full understanding of:

    - capital and operating cost implications
    - manpower staffing needs
    - maintenance requirements
    - nitrification performance
    - operating strategies

  The BAF process is an innovative process in North America and long-term operating experience is unavailable here. Operating facilities with a history of operation are located at:

    - North-Bierley, Britain
    - Lunéville, France
    - Soissons, France

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Mr. Mark Maki	City of Windsor
Mr. John Faust	City of Windsor
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Mr. Jack MacRae	City of Windsor
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Mr. Paul Robson	CH2M HILL ENGINEERING LTD.
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Section 1
INTRODUCTION

The City of Windsor has embarked on an important and costly program to upgrade its West Windsor Pollution Control Plant (PCP) from enhanced primary treatment to a form of secondary, biological treatment. The cost associated with this upgrade and the enhanced quality of the West Windsor PCP's current effluent has led the City to investigate alternative innovative treatment methods for this upgrade. The City is investigating these alternatives to reduce the upgrade's cost and to determine if alternative processes are suited to treating the existing plant's primary effluent.

The City determined that the plant upgrade will be required because the Ontario Ministry of the Environment (MOE) has indicated through the MISA program that Ontario wastewater treatment plants will be required to provide Best Available Treatment Economically Achievable (BATEA). These indications suggested that the West Windsor plant will require a form of secondary treatment, although the specific requirements have not yet been mandated by the MOE. Currently, twenty-four Ontario plants, which border on the Great Lakes, have the equivalent of primary treatment and will be affected by the MOE's expected requirements.

This report summarizes the first phase of the potential upgrade during which a number of treatment options were investigated. The study involved the operation of four pilot scale treatment processes using the existing plant effluent for an approximate one-year period. The results of these processes' operation, in conjunction with the expected capital and operation and maintenance (O&M) costs, will be used by the City to determine the best treatment process alternative for the West Windsor PCP upgrade. The City of Windsor retained CH2M HILL ENGINEERING LTD. in March 1990 to conduct this first phase study and to review alternative treatment processes for the eventual upgrade of the West Windsor PCP.

BACKGROUND

The West Windsor PCP is located on the west side of the City of Windsor and services approximately three-quarters of the City and the Town of LaSalle. The rated treatment capacity of the existing plant is 159,000 m^3/d (35 MIGD) with treatment capacity of peak flows to 350,000 m^3/d (77 MIGD) or a 2.2 times peaking factor (PF). Flows above the peak design flow receive screening, grit removal, and disinfection (seasonal) prior to discharge to the Detroit River. The existing West Windsor PCP is a physical-chemical wastewater treatment plant that provides enhanced primary treatment using chemical precipitation and coagulation and polymer addition for flocculation prior to primary sedimentation. The main treatment systems are outlined in Table 1.1. Dewatered sludge cake is currently lime-stabilized and is either hauled to a landfill to be used as daily cover or used on agricultural fields as a soil nutrient or conditioner.

Table 1.1
Main Treatment Systems at the West Windsor PCP

Process	Equipment/Item	Description
Influent Pumping Station	- six centrifugal pumps	- pump station designed with two wet-wells, with three pumps on each well - four pumps have variable-speed capability - four pumps rated at 370 kW, two pumps rated at 260 kW - maximum installed pumping capacity of 953,000 m^3/d
Bar Screens and Aerated Degrit Tanks	- two mechanically cleaned bar screens - six aerated degrit tanks	- 25 mm bar screen opening - 341,000 m^3/d capacity (each) - four tanks with 300 m^3 volume and hydraulic capacity of 112,500 m^3/d (each) - two tanks with 47 m^3 volume
Primary Clarification	- six primary clarifiers	- each 36.6 m in diameter, at 3.35 m SWD - each rated at 27,000 m^3/d average day capacity
Disinfection	- chlorination	- chlorine evaporators (2), chlorinators (3) and chlorine ejectors (2) - maximum dosage rate of 9 mg/L
Sludge Dewatering	- Solid bowl centrifuges	- two 112 kW units and one 93.3 kW unit - larger units rated at 2,250 L/min capacity, smaller unit rated at 375 L/min

The treated plant effluent has met MOE compliance limits for a primary facility (MOE Policy 08-01) for BOD_5, suspended solids (SS), and total phosphorus (TP) for the last three and one-half years. The compliance limits require an annual removal efficiency for total BOD_5 and SS of 50 and 70 percent, respectively, and a monthly average compliance for TP of 1 mg/L. Table 1.2 summarizes monthly performance data during the period from January 1988 to June 1991. One exception to consistent compliance was the excursion of TP to 1.1 mg/L in one month of this forty-two month period. The plant averaged 62 and 82 percent removal for total BOD_5 and SS.

Table 1.2
West Windsor PCP Historical Operational Data

Month	Treated Flow (1000 m3/d)	Monthly Average Concentrations (mg/L)					
		Influent			Effluent		
		TSS	TBOD5	TP	TSS	TBOD5	TP
Jan-88	111.0	99	84	3.4	19	33	0.6
Feb-88	119.0	108	79	3.4	27	26	0.8
Mar-88	107.0	109	91	3.7	31	34	0.8
Apr-88	117.0	75	69	3.3	28	33	0.6
May-88	102.0	148	111	4.7	35	38	0.9
Jun-88	102.0	115	97	4.2	31	40	0.7
Jul-88	107.0	141	101	4.1	22	31	0.4
Aug-88	118.0	103	72	3.9	26	25	0.6
Sep-88	119.0	105	80	3.9	28	31	0.5
Oct-88	121.0	114	106	3.8	27	39	0.6
Nov-88	142.0	97	79	3.2	28	35	0.6
Dec-88	110.0	152	106	4.3	30	40	0.9
Jan-89	109.0	95	85	3.9	24	33	0.6
Feb-89	103.0	128	101	4.9	20	41	0.7
Mar-89	112.0	151	104	4.3	26	34	0.7
Apr-89	132.0	104	89	3.2	19	35	0.4
May-89	113.0	183	130	4.6	23	43	0.9
Jun-89	146.2	118	73	3.7	20	29	0.5
Jul-89	133.3	95	66	3.0	23	29	0.5
Aug-89	118.5	99	68	3.6	25	30	0.7
Sep-89	130.6	81	85	3.5	23	44	0.7
Oct-89	114.0	124	98	3.8	26	57	0.8
Nov-89	117.6	130	112	4.0	28	44	0.6
Dec-89	90.5	180	132	5.0	34	50	1.1
Jan-90	127.4	147	114	3.5	26	42	0.7
Feb-90	163.1	125	77	2.8	24	35	0.7
Mar-90	135.0	147	85	3.5	19	29	0.6
Apr-90	129.5	138	90	3.4	19	29	0.6
May-90	138.8	179	113	4.4	16	50	0.6
Jun-90	126.3	179	118	4.2	13	48	0.5
Jul-90	111.9	172	85	4.6	14	26	0.4
Aug-90	129.9	147	86	3.6	16	25	0.5
Sep-90	162.8	133	81	3.5	16	24	0.6
Oct-90	155.7	132	97	3.7	19	32	0.7
Nov-90	126.6	157	111	4.5	25	36	0.8
Dec-90	162.6	148	121	4.0	34	39	0.8
Jan-91	127.9	140	112	3.8	23	39	0.6
Feb-91	117.5	125	117	3.6	25	38	0.6
Mar-91	115.9	143	115	3.8	25	36	0.6
Apr-91	138.1	131	96	3.4	25	38	0.6
May-91	128.7	153	116	3.4	35	51	0.8
Jun-91	130.8	92	92	3.3	25	31	0.5
Jul-91	110.2	137	110	4.5	25	41	0.7
Aug-91	119.0	113	89	3.2	24	28	0.6
Sep-91	105.0	143	99	4.1	25	37	0.7
Oct-91	118.5	165	148	3.8	30	46	0.8
Nov-91	113.4	146	137	4.1	27	36	0.6
Dec-91	111.7	168	126	3.6	21	35	0.4
Jan-92	115.8	203	166	4.4	19	41	0.4
Feb-92	143.0	145	124	3.3	23	37	0.5
Mar-92	144.8	153	114	3.6	20	31	0.5
Apr-92	151.6	148	108	3.5	18	28	0.4
Overall Period							
Average	**124.2**	**134**	**101**	**3.8**	**24**	**36**	**0.6**
Minimum	**90.5**	**75**	**66**	**2.8**	**13**	**24**	**0.4**
Maximum	**163.1**	**203**	**166**	**5.0**	**35**	**57**	**1.1**
Yearly Averages							
1988	**116.4**	**119**	**92**	**3.9**	**26**	**36**	**0.7**
1989	**118.3**	**124**	**95**	**4.0**	**24**	**39**	**0.7**
1990	**139.1**	**150**	**98**	**3.8**	**20**	**34**	**0.6**
1991	**119.7**	**138**	**113**	**3.7**	**26**	**38**	**0.6**
partial 1992	**138.8**	**162**	**128**	**3.7**	**20**	**34**	**0.4**

The West Windsor PCP obtains enhanced particulate removal in the existing plant by the addition of precipitant and polymer. The plant uses either alum or ferric chloride in conjunction with an anionic polymer. An investigation by the City regarding precipitant addition concluded that optimal dosage and removal efficiency is obtained by adding alum to the suction side of the raw sewage pumps or by adding ferric chloride to the effluent of the grit tank. The tank outlet has a hydraulic step that permits the ferric chloride to mix with the raw wastewater. Anionic polymer is added to the discharge of the two Montana flumes located in the channels upstream of the primary clarifiers. Table 1.3 summarizes precipitant and polymer usage from January 1990 to June 1991. (A site layout of the existing plant is shown in Section 4 of this report).

Table 1.3 shows that the plant on average uses 45,970 kg/mo of ferric chloride (as Fe^{+3}) or 19,280 kg/mo of alum (as Al^{+3}). The plant also uses an average of 1680 kg/mo of an anionic polymer in conjunction with the precipitant addition for enhanced particulate and phosphorus removal. This provides an average dosage of 12.1, 4.6, and 0.4 mg/L for Fe^{+3}, Al^{+3}, and polymer. MOE typical values for primary treatment are 16 mg/L for $FeCl_3$ (as Fe^{3+}) and 100 mg/L for alum (9.1 mg/L as Al^{+3}).

The plant produces 748 t/mo of dry sludge on average at 8.4 percent total solids. This sludge is dewatered prior to disposal. Typical MOE values for primary treatment with chemical precipitation for phosphorus removal are 3.2 L of sludge per cubic meter of flow treated at an average of 4.5 percent solids[21]. The West Windsor PCP currently treats an average flow of 4,090 ML/mo and based on the MOE typical values this would amount to 590 t/mo of dry sludge. Therefore, the plant produces more sludge than a typical primary plant with phosphorus removal due to the addition of polymer for enhanced settling. This results in a significantly higher quality effluent than would be expected based on MOE typical values without polymer addition. The West Windsor PCP also produces a sludge that is on average twice the concentration of a typical primary facility. The plant achieves this high primary sludge solids concentration using less than typical dosages of either $FeCl_3$ or alum in conjunction with anionic polymer.

Prior to the startup of the pilot scale treatment alternatives, input from the MOE was sought to determine the likely effluent requirements for an upgraded facility at West Windsor. The MOE indicated that, at the present time, the expected effluent design requirements for the upgraded plant would not be more stringent than 15, 15, and 0.5 mg/L for BOD_5, SS, and TP respectively. Nitrogen control may be a future objective. The MOE Windsor District Office has indicated that an effluent objective of 0.1 mg/L un-ionized NH_3-N may be required (see meeting notes, Appendix D). Effluent objectives are discussed further in Section 4. The proposed plant upgrade would include no treatment capacity increase above the current design rating. In addition, the current peaking factor of 2.2 would be maintained. Flows in excess of this flow are provided with screening, grit removal, and seasonal chlorination.

Table 1.3
West Windsor PCP Sludge Production and Chemical Usage

Date	Total Monthly Flow (ML/mo)	SS Concentration (mg/L)		Sludge Pumped (m3/mo)	Sludge Solids (% as TS)	Precipitant and Polymer Used (kg/mo)			Sludge Production (t/mo)	Dosage (mg/L)		
		Influent	Effluent			FeCl3 (as Fe+3)	Alum (as Al+3)	Polymer		FeCl3 (as Fe+3)	Alum (as Al+3)	Polymer
Jan-90	3949	147	26	7090	10.7	13726	19926	1500	761	3.5	5.0	0.4
Feb-90	4567	125	24	6829	10.7		23356	1600	730		5.1	0.4
Mar-90	4186	147	19	8690	7.7		21283	2575	671		5.1	0.6
Apr-90	3885	138	19	6473	8.2		17338	2125	531		4.5	0.5
May-90	4302	179	16	9059	8.2	30349	9626	2325	739	7.1	2.2	0.5
Jun-90	3788	179	13	10582	8.4		24015	1650	894		6.3	0.4
Jul-90	3469	172	14	11359	6.1		16550	1475	692		4.8	0.4
Aug-90	4027	147	16	10814	6.8		21349	1650	740		5.3	0.4
Sep-90	4884	133	16	9034	7.4		15713	1825	664		3.2	0.4
Oct-90	4826	132	19	11283	6.9		20038	1800	777		4.2	0.4
Nov-90	3797	157	25	13558	6.2		16550	1350	846		4.4	0.4
Dec-90	5041	148	34	10149	8.2		16637	1525	832		3.3	0.3
Jan-91	3964	140	23	7775	9.4	51317	2514	1250	727	12.9	0.6	0.3
Feb-91	3289	125	25	7058	9.1	44434		1095	643	13.5		0.3
Mar-91	3593	143	25	6522	10.2	40014		1385	666	11.1		0.4
Apr-91	4143	131	25	5853	10.5	44532		1500	615	10.7		0.4
May-91	3989	153	35	11192	9.3	48142		1375	1045	12.1		0.3
Jun-91	3925	92	25	12949	6.9	47373		2300	894	12.1		0.6
Average	**4090**	**144**	**22**	**9237**	**8.4**	**45969**	**19283**	**1684**	**748**	**12.1**	**4.6**	**0.4**
Maximum	**5041**	**179**	**35**	**13558**	**11**	**51317**	**24015**	**2575**	**1045**	**14**	**6**	**1**

Note : Amount of Ferric and Alum used and dosage based on months of complete usage

PROJECT OBJECTIVES

The objective of this study was to evaluate selected treatment alternatives at pilot scale for the future upgrade of the West Windsor PCP. The testing of each alternative was conducted to investigate each process' applicability for treating the existing effluent from the West Windsor PCP to obtain the target effluent requirements for the future upgraded plant. Specific objectives of the project were:

- Evaluate at pilot scale a maximum of four alternative processes capable of providing secondary treatment of the West Windsor PCP.
- Use the performance of the existing physical/chemical treatment facility to the maximum degree possible
- Determine potential effluent requirements for the future plant in consultation with the MOE Windsor District Office
- Determine organic and hydraulic loading criteria for each process, which would result in the process meeting the expected effluent requirements of the future upgraded plant
- Communicate ongoing results through progress reports and Steering Committee meetings
- Communicate project findings to LaFontaine, Cowie, Buratto and Associates Ltd. (LCBA) for their input in costing the alternatives
- Generate a report to summarize and document the performance of the treatment alternatives tested.

STEERING COMMITTEE

A Steering Committee was established by the City to review the results of the work in progress and to make recommendations on further work and pilot plant operational changes. The Steering Committee included members of the City of Windsor, MOE, Environment Canada, the University of Windsor, LaFontaine, Cowie, Buratto and Associates Ltd., and CH2M HILL ENGINEERING LTD. The Regional MOE office was kept informed of the study progress and results by the MOE District office representative on the Steering Committee. The Steering Committee representatives were:

City of Windsor, Department of Public Works

Mr. L. S. Romano, P.Eng., Director of Pollution Control and Assistant Commissioner of Works
Mr. J. Faust, B.Sc., Head Chemist
Mr. M. Maki, B.Sc., West Windsor PCP Superintendent
Dr. R. Drynan, P.Eng., Senior Environmental Engineer
Mr. J. MacRae, B.Sc., Laboratory Staff

Ministry of the Environment

Mr. H. Monteith, P.Eng., Senior Project Engineer, Water Resources Branch
Mr. A. K. Ho, P.Eng., Supervisor STP Compliance Unit, Water Resources Branch
Mr. A. B. Braganza, P.Eng., Senior Specialist, Approvals Branch
Ms. G. Griffore, Environmental Officer, Windsor District Office
Mr. A. O. Stephens, P.Eng., Senior Environmental Officer, Windsor District Office

University of Windsor

Dr. J. K. Bewtra, P.Eng., Professor, Department of Civil Engineering, University of Windsor

Environment Canada

Mr. G. Speirs, M.Sc., Project Engineer, Environment Canada/Wastewater Technology Centre

LaFontaine, Cowie, Buratto and Associates Ltd.

Mr. H. S. Horneck, P.Eng., Director

CH2M HILL ENGINEERING LTD.

Mr. J. P. Stephenson, P.Eng., Department Manager
Mr. M. L. Newbigging, P.Eng., Project Engineer
Mr. P. Robson, CET, Senior Field Technologist

Mr. H. Monteith represented the Water Resources Branch until March 1991; Mr. T. Ho became the Water Resources Branch representative in March 1991. Mr. A. Braganza attended the Steering Committee meetings after May 1991, but was informed of the project performance prior to that point. Mr. G. Speirs began attending Steering Committee meetings in February 1991.

PROJECT AND REPORT FORMAT

CH2M HILL ENGINEERING LTD. was retained by the City to conduct the first phase of the potential West Windsor PCP upgrade in March 1990. The project was initiated with mobilization meetings held on March 19, 1990 and April 11, 1990. During these meetings, the processes to be tested and the target effluent concentrations were discussed and finalized. The pilot plant units were constructed in the beginning of May 1990. The pilot units began operation in late June 1990 and the last unit was operational by August 1990. All pilot plant units were operated until the end of May 1991. The TF/SC and BAF processes were operated for an initial extension period (June 1, 1991 to August 6, 1991) and a follow-up extension period (September 17, 1991 to April 24, 1992) to resolve outstanding issues.

Nine progress reports were generated during the first year and three during the extension period and distributed to the committee members. Each progress report summarized the pilot plant performance over the period from the last progress report including effluent performance, loading parameters, and operations. A copy of each progress report, excluding the raw data, is provided in Appendix A. Appendix A also contains a chronological list of milestones for each process. Table 1.4 summarizes the meeting dates and deliverables to the Steering Committee during the study period.

Table 1.4
Timetable of Meeting Dates and Project Deliverables

Date	Meeting No. and/or Deliverable
March 19, 1990	• Meeting No. 1, kick-off
April 11, 1990	• Meeting No. 2, mobilization meeting
May 7, 1990	• Meeting No. 3 • Technical Memorandum No. 1 delivered to City, discussing Treatment Alternatives for West Windsor • Meeting held between City and CH2M HILL on May 16, 1991 to discuss pilot plant construction and requirements
August 22, 1990	• Meeting No. 4 • Memorandum discussing site visits to Trickling Filter/Solids Contact (TF/SC), Rotating Biological Contactor (RBC) and Biological Aerated Filter (BAF) plants delivered • Progress Report No. 1 delivered • All pilot plants operating
Sept. 25, 1990	• Meeting No. 5 • Progress Report No. 2 delivered
Oct. 30, 1990	• Meeting No. 6 • Progress Report No. 3 delivered
November 28, 1990	• Meeting No. 7 • Progress Report No. 4 delivered • Meeting held between City and CH2M HILL on Dec. 6, 1990 to discuss TF/SC loading and winterization of pilot plants
December 18, 1990	• Meeting No. 8 • Progress Report No. 5 delivered
January 17, 1991	• Meeting No. 9 • Progress Report No. 6 delivered
Feb. 19, 1991	• Meeting No. 10 • Progress Report No. 7 delivered
March 18, 1991	• Meeting No. 11
April 11, 1991	• Meeting No. 12 • Progress Report No. 8 delivered
May 7, 1991	• Meeting No. 13 • Progress Report No. 9 delivered
May 15, 1991	• Meeting No. 14
May 29, 1991	• Technology Transfer Seminar (May 29, 1991) conducted by City of Windsor and MOE, presentations by the City, MOE, CH2M HILL and selected equipment suppliers
June 4, 1991	• Meeting No. 15 • ASP and RBC pilot plants operated by City staff until June 14, 1991 • BAF and TF/SC operated by City staff until August 6, 1991 • Environment Canada to provide emissions date.
July 17, 1991	• Meeting No. 16
August 13, 1991	• Meeting No. 17 • BAF and TF/SC pilot plant shutdown on August 6, 1991 • Interim meeting held on August 26, 1991 to discuss comparison approach to evaluate each process
August 26, 1991	• Meeting No. 18, Evaluation Procedure
October 3, 1991	• Meeting No. 19, Extension Period Performance
October 15, 1991	• Meeting No. 20, Extension Period Performance

Table 1.4
Timetable of Meeting Dates and Project Deliverables

Date	Meeting No. and/or Deliverable
November 4, 1991	• Meeting No. 21 • Review Draft Final Report with Steering Committee
November 20, 1991	• Meeting No. 22 - Extension Period Performance • Presentation by John Meunier • Progress Report E.1 delivered (November 11, 1991)
January 8, 1992	• Meeting No. 23 - Extension Period Performance • Progress Report No. E-2 sent out after meeting (January 16, 1992)
March 3, 1992	• Meeting No. 24 - Extension Period Performance • Presentation by John Meunier
June 23, 1992	• Meeting No. 25 - Extension Period Performance • Progress Report E-3 delivered
October 15, 1992	• Meeting No. 26 • Review revised section of CH2M HILL report and costing by LCBA
November 20, 1992	• Final Report Submission
November 26, 1992	• Meeting No. 27 • Review Final Report with Steering Committee

This report is presented in six sections. These sections are:

Section 1	Introduction	Introduction of the project, background associated with the West Windsor PCP and upgrading requirements, project and report format
Section 2	Method	Outlines the process technologies tested, pilot plant specifications, pilot testing schedule and analytical and monitoring sampling schedule
Section 3	Results of Pilot Plant Operations	This section provides the results of the pilot plant operations in terms of effluent performance, hydraulic and organic loadings, sludge handling and operation and maintenance. A separate section is provided for each process.
Section 4	Full-Scale Facility Process Implications	This section reviews preliminary implications of each unit in terms of a full-scale facility at the West Windsor PCP, based on pilot-plant operations and site inspections of full-scale plants. This section also summarizes the costing full-scale cost estimates for each process.

Section 5	Overall Comparison of Process Alternatives	This section summarizes study results and comparatively evaluates each alternative.
Section 6	Conclusions and Recommendations	Conclusions and recommendations of study
Section 7	References	
Section 8	Glossary of Terms	

Section 2
METHOD

This section discusses the processes which were pilot tested, the testing schedule, and sampling and monitoring requirements. A review of each pilot plant process was initiated to determine loading and operating parameters.

After discussions between the City and CH2M HILL, a number of treatment alternatives were selected for consideration. Figure 2-1 depicts a block diagram of the treatment alternatives selected for consideration for the West Windsor PCP upgrade. The list of potential alternatives was reduced to evaluate distinct processes (i.e. suspended growth, hybrid and fixed film) and to obtain a maximum of four units for testing. The four processes selected were:

- Trickling Filter/ Solids Contact Process (TF/SC)
- Rotating Biological Contactor Process (RBC)
- Biological Aerated Filter Process (BAF)
- Activated Sludge Process (ASP)

The process alternatives and the method used to conduct the pilot plant testing are discussed below.

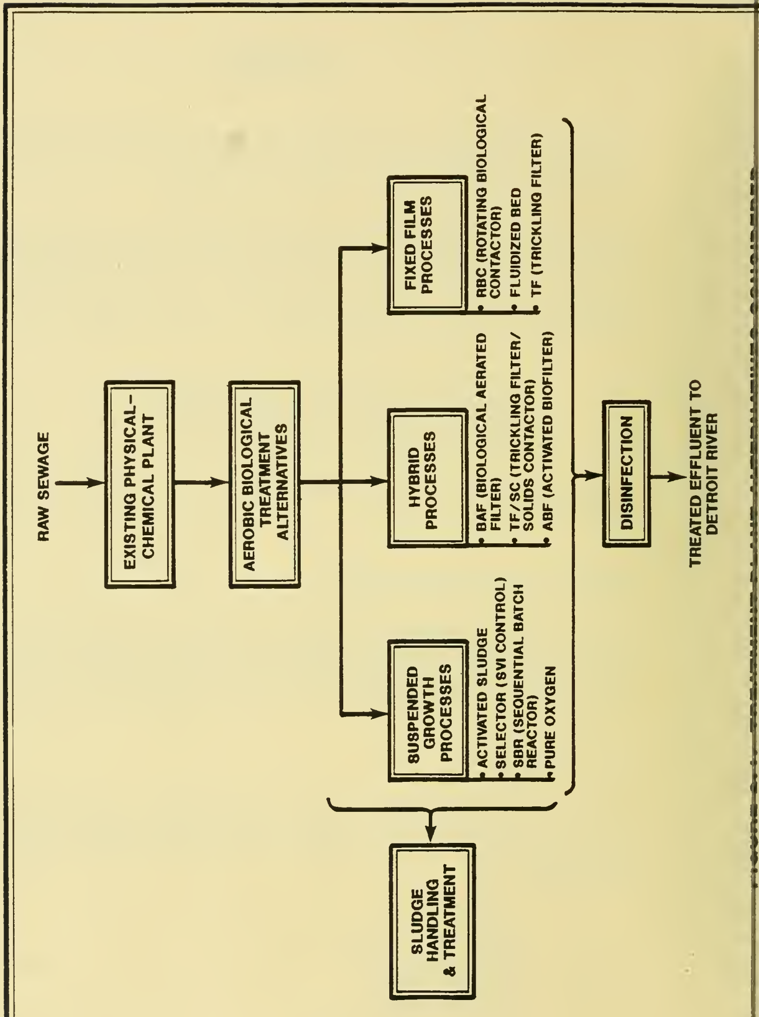
REVIEW OF TECHNOLOGIES

Prior to designing and sizing the pilot plants, a review of each technology was initiated. The results of this review were submitted to the Steering Committee as a Technical Memorandum in May 1990 (Appendix B). This section reviews the main aspects of each process and presents additional information obtained from other sources during the study.

TF/SC PROCESS

The use of trickling filters with rock media was at its height in the United States during the 1940s. At that time nearly 60 percent of the secondary wastewater treatment plants in the United States were trickling filters[1]. The main advantages associated with the trickling filter were:

- simplicity
- thick secondary sludge
- low operating costs
- shock load resistance
- low maintenance
- low power requirements


RAW SEWAGE
EXISTING PHYSICAL–CHEMICAL PLANT
AEROBIC BIOLOGICAL TREATMENT ALTERNATIVES
SUSPENDED GROWTH PROCESSES
• ACTIVATED SLUDGE
• SELECTOR (SVI CONTROL)
• SBR (SEQUENTIAL BATCH REACTOR)
• PURE OXYGEN
HYBRID PROCESSES
• BAF (BIOLOGICAL AERATED FILTER)
• TF/SC (TRICKLING FILTER/SOLIDS CONTACTOR)
• ABF (ACTIVATED BIOFILTER)
FIXED FILM PROCESSES
• RBC (ROTATING BIOLOGICAL CONTACTOR)
• FLUIDIZED BED
• TF (TRICKLING FILTER)
SLUDGE HANDLING & TREATMENT
DISINFECTION
TREATED EFFLUENT TO DETROIT RIVER

Activated sludge plants became more popular mainly due to the trickling filter's higher initial capital cost and the need to meet more stringent effluent requirements.

A number of advances have occurred since the introduction of the trickling filter that make it a potential candidate for a secondary facility with today's increasingly more stringent effluent requirements. These advances include the development of better media and the discovery of new combined processes.

The development of new media including vertical filter media, horizontal filter media, random media, and cross flow media has increased the void area available compared to rock media. These new types of media provide increased air circulation, more surface area for biological growth and the ability to slough growth without plugging. Also, combined processes have improved the effluent quality attainable from the trickling filter process alone. These processes include:

- trickling filter solids contact (TF/SC)
- activated biofilter (ABF)
- roughing filter activated sludge (RF/AS)
- biofilter activated sludge (BF/AS)
- trickling filter activated sludge (TF/AS)
- two-stage trickling filter (TF/TF)
- trickling filter/trickling filter/solids contact (TF/TF/SC)

Operational problems associated with a trickling filter include:

- odours
- ponding on filter media
- icing
- filter flies
- snail shell accumulation

The TF/SC process was proposed by the City for the West Windsor PCP because of the low influent concentration that would originate from the existing enhanced primary plant. The trickling filter can be designed to obtain organic carbon removal and nitrification; the solids contactor is designed primarily to condition the trickling filter effluent floc to enhance settling. The solids contactor is also expected to provide additional organic removal and nitrification.

Data from four plants[2] have shown that a TF/SC plant can obtain effluent quality between 7 and 21 mg/L for total BOD_5 and between 8 and 13 mg/L for SS depending on the loading and type of trickling filter media. This effluent quality was obtained at a total organic loading rate (TOL) varying from 0.4 to 1.8 kg total $BOD_5/m^3 \cdot d$, a solids contact and sludge re-aeration time of 9 to 39 minutes, and a secondary clarifier overflow rate of 12 to 29 $m^3/m^2 \cdot d$. Other references[1,3] indicate similar loadings of 0.5 kg $BOD_5/m^3 \cdot d$ and 0.64 kg $BOD_5/m^3 \cdot d$ for TOL, 30 min and 60 min solids contact time, and 19 $m^3/m^2 \cdot d$ secondary clarifier overflow rate. Most references also discuss a wetting rate on the trickling filter surface area. These rates[1,3] vary from 0.5 to 0.7 $L/m^2 \cdot s$ (43 to 60 $m^3/m^2 \cdot d$).

For a single stage trickling filter to nitrify, a reduced TOL and wetting rate may be required. One media supplier[4] suggests that the wetting rate be maintained at no higher than 0.34 $L/m^2 \cdot s$ (29 $m^3/m^2 \cdot d$) at a wastewater temperature between 7 and 18°C to obtain 90 percent removal of NH_3-N. Work conducted at two plants in Texas[5] suggests that a TOL of 0.4 kg $BOD_5/m^3 \cdot d$ is the maximum loading to obtain a 80 percent nitrification in the trickling filter. Wastewater temperatures vary from a high of 20°C to a low of 14°C at these plants.

RBC PROCESS

Development of the rotating biological contactor (RBC) process can be traced back to the 1920s in both Germany and the United States. Initial units were constructed with wooden disks and proved to be impractical to manufacture and prone to deterioration. In the 1950s, experimentation began with plastic disks which began to make the process viable. Units still had a higher capital cost, but required low maintenance and energy consumption. The construction of more compact disks, in 1972, with more surface area for a given volume, was an important development[7].

Some advantages of the RBC process are:

- ability to produce high quality effluent
- small land requirements
- resistance to washout at high loadings (organic and hydraulic)
- small hydraulic grade-line
- low energy consumption
- simple operation

Although mechanical problems have been eliminated or minimized in recent years[3], the RBC process has also been associated with a number of disadvantages, including:

- structural integrity problems (shaft failures and media deterioration)
- requirement for enclosures
- need for good grit and primary clarification upstream
- problems with filamentous growth (white biomass, Thiothrix or Beggiatoa organisms)
- odours

RBC plant operation depends on the organic and hydraulic loading rates. RBC plants that obtain effluent quality of 10 mg/L total BOD_5 and 5 mg/L NH_3-N are common[6,7,8]. Plants are generally designed based on an organic and ammonia loading rate per unit surface area of media and a hydraulic loading rate. A range of reported organic loading rates vary from 1 to 29 g $BOD_5/m^2 \cdot d$[6,7,8]. Hydraulic loading rates vary from 40 to 244 $L/m^2 \cdot d$[10]. An example of combined organic carbon removal and nitrification is documented for a 22,700 m^3/d facility[7] serving a major U.S. army installation. This facility was designed to meet a 10 mg/L total BOD_5 and 5 mg/L NH_3-N effluent at an organic loading of 4.4 g $BOD_5/m^2 \cdot d$ and 1.0 g $NH_3\text{-}N/m^2 \cdot d$. Combined organic carbon removal and nitrification in an RBC plant is often expressed as only an organic loading. Manufacturers[8] have reported loadings between 8 and 17 g $BOD_5/m^2 \cdot d$ for combined BOD_5 removal plus 90 percent nitrification. Extensive pilot testing[10] has shown that organic loadings between 6.0 to 6.5 g $BOD_5/m^2 \cdot d$ can obtain an average effluent objective of 15 mg/L for BOD_5 and SS at a wastewater temperature of 10°C. To obtain nitrification to less than 5 mg/L TKN, at similar winter wastewater temperatures, a maximum nitrogen loading of 0.4 g $NH_3\text{-}N/m^2 \cdot d$ is recommended.[10]

Final clarification surface overflow rate (SOR) for an RBC process is recommended to be 16 to 20 $m^3/m^2 \cdot d$ to meet effluent requirements between 10 to 15 mg/L SS[9]. Precipitate addition may also be required to obtain effluent SS concentrations between 10 to 15 mg/L.

BAF PROCESS

Recently, high-rate biological processes that reduce reactor volume and land area requirements as well as provide good effluent quality have been developed. An upflow fluidized bed reactor with pre-oxygenation, an upflow granular media reactor with external air addition, an upflow plastic media with external air, and a downflow granular media reactor with external air addition are examples of newly developed units[11].

The downflow granular media reactor was first awarded a patent in 1978 following preliminary laboratory and full-scale development of the technology by OTV of France. The process has since been used in many European wastewater treatment facilities and is considered an innovative secondary treatment process alternative in North America. Five full-scale U.S installations ranging in size from 750 to 7,570 m^3/d are presently in service[11]. In Canada, there are presently three units in operation in Quebec, and at least two more facilities are in the design phase[12]. The largest plant in Quebec began operation in May 1991 and is located in Sherbrooke. The plant is rated at 80,000 m^3/d hydraulic capacity for a 120,000 person equivalent, 7,470 kg BOD_5/d, 8,140 kg SS/d, and 400 kg TP/d loading. Effluent requirements are 20, 20, and 0.8 mg/L for BOD_5, SS, and TP without a nitrification requirement.

The BAF process consists of a granular media filter, generally of expanded shale, through which pre-treated wastewater is passed. A number of nozzles pass through an underdrain plate that supports the media. Process air is supplied to the bed about 25 cm above the underdrain plate. The process provides removal of organic carbon, suspended solids, and nitrification in a single reactor. Accumulated solids and biomass are stored in the media and removed by periodic backwashing. Backwashing can be set at a predetermined schedule, generally once a day, or based on the headloss through the media.

BAF plant designs are based on organic and hydraulic loading criteria. If nitrification is required, reduced organic and hydraulic loadings are recommended. One source indicated that an organic loading varying from 1 to 3 kg total $BOD_5/m^3\cdot d$ would result in an effluent BOD_5 concentration of less than 15 mg/L with a wastewater temperature varying from 10°C to 16°C[13]. The same source showed results for combined organic carbon removal and nitrification at loadings up to 2 kg total $BOD_5/m^3\cdot d$ and 0.3 kg $NH_3\text{-}N/m^3\cdot d$. Effluent ammonia concentrations were generally less than 6 mg/L during this study at wastewater temperatures as low as 10°C. Pilot studies showed that effluent BOD_5 concentrations less than 15 mg/L could be obtained at an organic loading up to 15 kg $COD/m^3\cdot d$[11]. Nitrification was also evident at these loading rates; however, nitrification was inhibited as the organic loading increased to 4.5 kg $COD/m^3\cdot d$. Nitrification was mainly governed by the nitrogen loading. Nitrification varied from 90 to 84 percent as the nitrogen loading increased from 0.16 to 0.59 kg $TKN/m^3\cdot d$. Results from three treatment plants indicate hydraulic loadings varying from 0.7 to 1.8 $m^3/m^2\cdot h$[11]. One of these plants was loaded at an average loading of 1.7 kg $BOD_5/m^3\cdot d$, 0.3 kg $NH_3\text{-}N/m^3\cdot d$, and a hydraulic loading of 0.7 $m^3/m^2\cdot h$ and obtained effluent BOD_5 and SS concentrations of 10 mg/L and partial nitrification.

The pilot plant equipment supplier indicated the following loadings for organic carbon removal at a wastewater temperature of 15°C[12]:

- an organic loading of 6 kg $BOD_5/m^3\cdot d$
- a hydraulic loading to a maximum of 7 $m^3/m^2\cdot h$

For combined organic carbon removal and nitrification, the supplier suggested loadings at a wastewater temperature of 15°C:

- up to an organic loading of 2.2 kg $BOD_5/m^3\cdot d$,
- hydraulic loadings to 2.2 $m^3/m^2\cdot h$
- a nitrogen loading of 0.7 $TKN/m^3\cdot d$

The supplier's loading estimates were based on the primary effluent characteristics from the West Windsor PCP. The effect of wastewater temperature on the ability of the BAF to nitrify has been investigated[14]. These results indicate that a maximum loading of 0.5 kg $NH_3\text{-}N/m^3\cdot d$ be used to obtain 90 percent removal at a wastewater temperature of 6°C.

ASP PROCESS

The activated sludge process was developed by Fowler et al. at the Manchester, England treatment plant in the early 1900s[3]. Several installations were put into operation in Canada and the United States in the 1920s and the process became widespread in the 1940s. Currently, the activated sludge process is the most common secondary treatment process in Canada.

The City of Windsor is familiar with the activated sludge process from the operation of its Little River PCP. The Little River PCP is an activated sludge plant hydraulically rated at 63,700 m^3/d. The plant obtains excellent organic carbon and particulate removal as well as essentially complete nitrification. A study[15] was conducted at the Little River PCP Plant No. 1 (36,400 m^3/d) in 1989 that showed that the plant could meet its MOE Certificate of Approval effluent compliance limits of 15, 15, 0.5, and 6 mg/L for total BOD_5, SS, TP, and NH_3-N at hydraulic loading rates which were higher than MOE design guidelines[21]. Table 2.1 shows the performance of the Little River PCP during the three season evaluation compared to MOE design guidelines.

PILOT PLANT OPERATIONAL CRITERIA

The pilot plants were located on the plant property and received the same influent characteristics using a portion of the effluent from the existing plant. All four pilot plants were operated from June 1990 to June 1991. The TF/SC and BAF pilot plants were operated for an additional year (July 1991 to April 1992) to resolve outstanding issues. The four units were located on the property of the West Windsor PCP, in an unused compost building. The building was open on one side; therefore, the units were operated at actual atmospheric conditions but sheltered from wind and direct sunlight. Effluent from two of the six primary clarifiers was pumped into a common feed well. Each of the four pilot plants were fed from this central tank or feed well; the volume of the feed well was 3.9 m^3. The well had an influent flow rate of approximately 680 L/min with a 100 mm overflow pipe to drain. The flow rate to the well was maintained by a submersible pump that ran continuously and was located in the primary effluent channel for primary clarifiers No. 5 and 6. Each of the pilot plants (BAF, RBC, TF/SC) had its own speed-controlled centrifugal pump, drawing primary effluent from the main feed well. The ASP feed was supplied by syphon from the main feed well to its own feed well where it was drawn off by a peristaltic pump. The feed pumps for the BAF, RBC, and TF/SC were protected by low-level floats. The maximum draw-down of the feed well was approximately 400 L/min (i.e. peak flows or storm flow conditions), leaving an overflow rate of 280 L/min. Each of the pilot plants received homogenous feed because the filling rate of the feed well and the angle at the point of entry created a well-mixed tank. Figure 2-2 is a schematic layout of the pilot plants indicating the pumping points and the main process reactors.

The sizing of the pilot plants is discussed in the Technical Memorandum dated May 4, 1991 (Appendix B). Key items from this memorandum and additional sizing or requirement changes that occurred during the pilot testing are summarized below.

Table 2.1
Summary of Performance of Little River PCP During Stressing of 1989

Parameter	Winter	Spring	Summer/Fall	MOE Guideline or Other
Average Daily Total Flow (10^3 m^3/d)	21.2 (4.7 MIGD)	25.4 (5.6 MIGD)	23.6 (5.2 MIGD)	-
Peak Daily Flow (10^3 m^3/d)	26.5 (5.8 MIGD)	37.0 (8.1 MIGD)	34.1 (7.5 MIGD)	-
Aeration Tank				
- HRT (h)	3	2.3	2.5	6
- Organic Loading (kg $BOD_5/m^3 \cdot d$)	0.64	0.53	0.42	0.3 - 0.71
- Oxygen Requirement (kg O_2/d)	1209 (carb) 2452 (carb+nitr) 2219 (C,N+D)*	927 (carb) 2377 (carb+nitr) 2248 (C,N+D)*	850 (carb) 2170 (carb+nitr) 2040 (C,N+D)*	2021 @ 2 mg/L DO 2455 @ 0 mg/L DO
- Dissolved Oxygen (mg/L)	2.4 (No. 1) 3.1 (No. 2)	2.2 (No. 1) 3.6 (No. 2)	1.3 (No. 1) 5.3 (No. 2)	2 mg/L
- F/M Ratio (d^{-1})	0.35	0.29	0.28	0.05 - 0.25 (nitr) 0.25 - 0.5 (carb)
- SRT (d)	7.7	6.6	6.1	>4 @ 20°C >10 @ 5°C (nitr)
Final Clarifiers				
- Peak SOR ($m^3/m^2 \cdot d$)	28.3	39.2	36.5	29.4 (nitr)
- Peak SLR ($kg/m^2 \cdot d$) @ 2500 mg/L MLSS	105	142	102	120 (@ 100% recycle)
- RAS Flow (% of influent flow)	49.0	46	44	100%

*Indicates oxygenation requirement for carbonaceous and nitrogenous removal with denitrification.

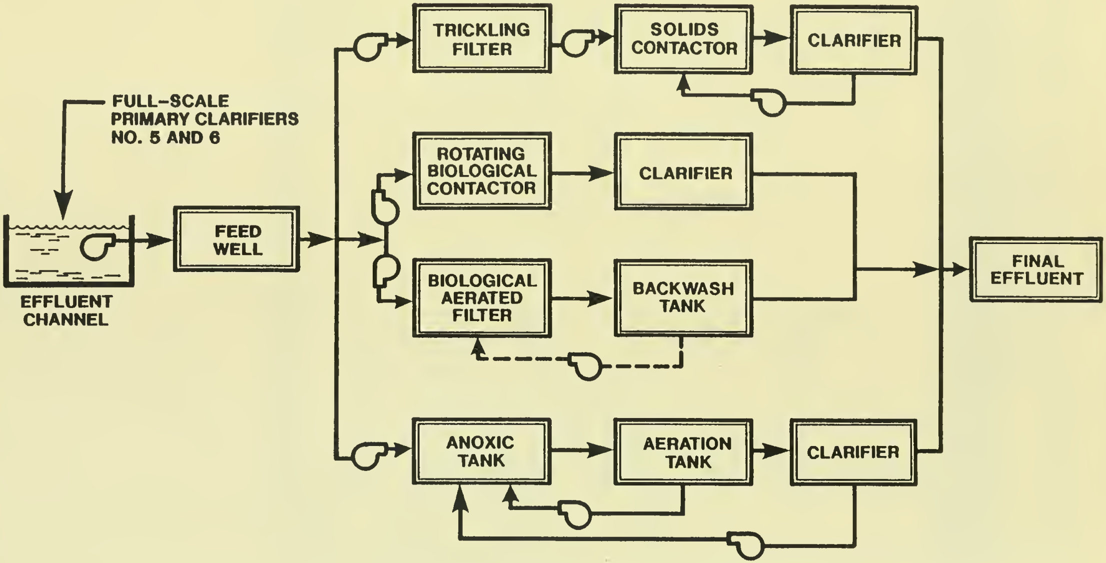


FIGURE 2.2: WEST WINDSOR PILOT PLANT FACILITIES

Table 2.2
TF/SC Pilot Plant Loading Characteristics

Parameter	Pilot Plant	Guideline
General		
• Flow (m^3/d)		
Avg.	175	
Peak Day	385	
• Influent Conc. (mg/L)		
$TBOD_5$	36	
TSS	26	
TKN	14.3	
NH_3-N	8.9	
TP	0.7	
Trickling Filter		
• Volume (m^3)	25.4	
• Surface Area (m^2)	5.9	
• TOL $\left(\frac{kgBOD_5}{m^3 \cdot d}\right)$	0.25	0.64 - 0.96
• Wetting Rate ($m^3/m^2 \cdot d$)	29.4	29.4
Solids Contact Tank		
• Volume (m^3)	4.5	
• HRT (min) (recycle included at .25 treated flow)	30	< 60
• Recycle Rate (m^3/d)	44	25% of treated flow
• SRT (d) ($Q_{WAS} = 2.3\ m^3/d$)	2	2
Clarifier		
• Surface overflow rate ($m^3/m^2 \cdot d$)	37.5	16

Table 2.3
RBC Pilot Plant Loading Characteristics

Parameter	Pilot Plant	Guideline
General		
- Flow (m^3/d)		
Avg.	65	
Peak Day	143	
- Influent Conc. (mg/L)		
$TBOD_5$	36	
TSS	26	
TKN	14.3	
NH_3-N	8.9	
TP	0.7	
RBC		
- Disc Diameter (m)	2	
- Disc Surface Area (m^2)	734	
- Organic Load (g $BOD_5/m^2 \bullet d$)	3.2	12-20 (soluble)
- Nitrogen Loading (g $TKN/m^2 \bullet d$)	1.3	0.3-1.1
- Hydraulic Loading ($L/m^2 \bullet d$)	88	40-244 (90 $L/m^2 \cdot d$ based on Environment Canada Study)
Clarifier		
- SOR ($m^3/m^2 \bullet d$)	32.5	16-20

Table 2.4
BAF (Biocarbone) Loading Characteristics

Parameter	Pilot Plant	Guideline
General		
• Flow (m^3/d)		
Avg.	30	
Peak Day	66	
• Influent Conc. (mg/L)		
$TBOD_5$	36	
TSS	26	
TKN	14.3	
NH_3-N	8.9	
TP	0.7	
BAF		
• Surface Area (m^2)	0.5	
• Volume (m^3)	1	
• Hydraulic Loading ($m^3/m^2\bullet h$)	2.5	2.0
• Organic Loading (kg $BOD_5/m^3\bullet d$)	1.1	2.2
• Nitrogen Loading ($kgTKN/m^3\bullet d$)	0.5	<0.7
• HRT (min) (based on empty bed)	48	30 - 60

Table 2.5
Bench-Scale Activated Sludge Process Loading Characteristics

Parameter	Pilot Plant	Guideline
General		
• Flow (mL/min)		
Avg.	70	
Peak Day	154	
• Influent Conc. (mg/L)		
$TBOD_5$	36	
TSS	26	
TKN	14.3	
NH_3-N	8.9	
TP	0.7	
Anoxic Zone		
• Volume (L)	1.4	
• HRT (h)	0.33	
Aeration Tank		
• Volume (L)	10.5	
• HRT (h)	2.5	6
• Organic loading (kg $BOD_5/m^3 \cdot d$)	0.34	0.3 - 0.71
• F/M (d^{-1})	0.18	0.3
Clarifier		
• Peak SOR ($m^3/m^2 \cdot d$)	7.5	29.4 (nitr)

TF/SC PROCESS

Based on the review of a number of processes, the process loadings shown in Table 2.2 were used to start up the pilot plant at the West Windsor PCP. To obtain accurate results from the pilot plant, a full-depth TF was used, and the surface area was made sufficiently large to minimize wall effects. The media used was Munters cross flow media, Blodek 19060 with a specific surface area of 138 m^2/m^3.

The influent was pumped from the main feed well to the top of the TF tower using a variable speed pump. The flow was evenly distributed across the filter surface area and collected in a tank at the bottom of the TF.

Located above the top of the media and distribution system in the tower, was a 500 L syphon tank where the filling and subsequent syphoning simulated the rotation of the distribution arms found in a full-scale plant TF. The syphon tank was taken out-of-service in December 1990 to facilitate bypassing after the SC tank. When the syphon engaged, the primary effluent left the tank through two 75 mm PVC pipes to a common trough that ran the width of the tower. Eight 50 mm PVC pipes that extended to the outer walls of the tower ran perpendicular to the trough. Each of the sixteen 50 mm PVC pipes contained four distribution heads: there were sixty-four heads in total or one distribution head for each square foot of media at the top of the tower.

The TF tower was started on July 6, 1990, with the effluent discharged directly to the drain. The solids contactor basin was completed July 19, 1990 and the TF/SC process was fully started on this date. The TF effluent was directed into a tank where it was pumped up to the first of two solids contactor basins via a low shear Hidrostal pump that was protected by low-level floats.

The solids contactor consisted of two circular tanks that could be operated in series or individually. Each tank contained a volume of approximately 2,000 L and was aerated by eleven fine bubble domes diffusers. The aerated effluent flowed by gravity from the second basin through a standpipe to the centre ring of the 2.4 diameter final clarifier. The settled solids from the clarifier were returned to the first solids contactor through a centrifugal pump.

Initially, a mixer was installed in the SC tank to prevent mixing limitations; however, the diffused air flow rate and the smooth action of the fine bubble diffusers made the use of the mixer unnecessary. The mixer was moved to the TF effluent tank to prevent biomass settling. The mixed liquor suspended solids (MLSS) from the two SC tanks flowed by gravity to a secondary clarifier. Return activated sludge (RAS) was returned to the SC by a variable speed pump. Wasting occurred from the SC in batch mode or constantly from the RAS line. The pilot plant schematic is shown in Figure 2-3.

The surface overflow rate used for the final clarifier was initially at twice the guideline flow rate due to equipment availability. Since the TF and SC were separate processes, it was felt that flow diversion after the TF tower, prior to pumping, could be used if necessary to reduce the final clarifier hydraulic loading.

RBC PROCESS

Based on the review of a number of processes and especially pilot studies conducted by Environment Canada, the process loadings shown in Table 2.3 were used for the RBC pilot plant[8]. The Environment Canada pilot studies were applicable to this work since the same pilot plant unit was used. The RBC pilot plant was an Envirex mechanically driven four-stage unit, Model No. 3048.

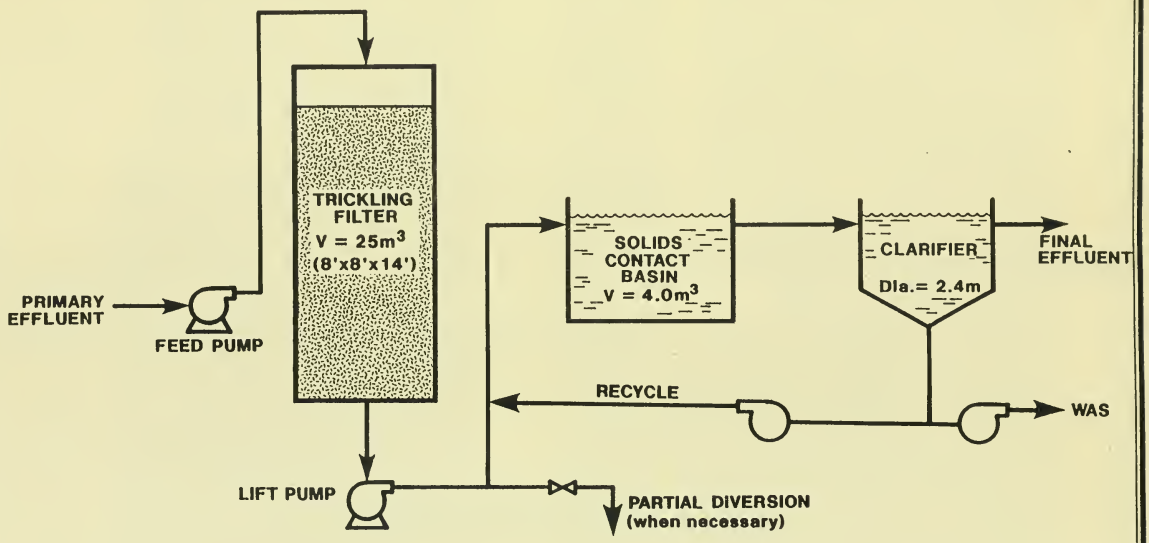


FIGURE 2.3: TF / SC PILOT PLANT

Influent to the four-stage RBC was pumped using a variable speed pump from the main feed well. Gravity flow was used through the RBC and final clarifier. The four-stage system used a serpentine-type flow, which is similar to a plug-flow set-up. The five access latches located on the cowling, were left open during startup and warm weather to allow any generated bio-gas to be ventilated and to provide additional ventilation. During the winter months, the hatches were left closed to minimize temperature drop over the exposed media. The clarifier waste pump was a progressing cavity type. Wasting on a continuous basis was not required due to the limited volume of sludge; therefore, the pump speed was set at approximately 8 - 9 L/min, with On/Off controlled by a variable timer. The original cycle for the wasting pump was 60 minutes "Off" and 2 minutes "On", 24 hours per day. This cycle created septic and floating sludge in the clarifier. A number of different cycles were tried to alleviate the problem and minimize the wasting volume. It was concluded that an "Off" time of 15 minutes and an "On" time of 1 minute, and a pumping rate of approximate 8.5 L/min was best suited to this pilot. This resulted in the pump being cycled on 96 times per day with a total waste volume of approximately 820 L/d. No attempt was made to maximize the thickness of the RBC sludge.

The equipment sizes and requirements for the pilot plant are shown in Figure 2-4. The initial clarifier surface overflow rate used on the pilot plant was twice the average day guideline loading. Clarifier availability made it expedient to begin with this loading with the provision to divert a portion of the RBC effluent if the clarifier appeared to be overloaded.

BAF PROCESS

The pilot plant used for pilot testing at the West Windsor PCP was leased by the City from J. Meunier Inc. of Montreal. The supplier, J. Meunier, refers to the process as the BioCarbone process. The unit loadings were based primarily on those values provided by the supplier; these values are shown in Table 2.4. Influent to the BAF was pumped using a variable speed pump. Controls and automatic valves were included to shut off the feed pump when the unit was backwashed. Flow was distributed to the submerged media bed by a splash plate. Effluent was collected in the bottom of the filter under the media through nozzles placed in the supporting plate. Submergence of the bed was controlled by a stand pipe located in the backwash tank. Overflow from the backwash tank flowed by gravity to drain. Once a day, or when filter clogging was observed, a backwash event was initiated. Filter clogging was based on a level electrode indicating the limiting headloss through the unit. Each backwash included five to six series of cleaning cycles. Each cleaning cycle included air and air with water backwashing to remove the accumulated biological particulates growth. Backwash water was wasted to drain by engaging a syphon. The equipment sizes and requirements for the pilot plant are shown in Figure 2-5.

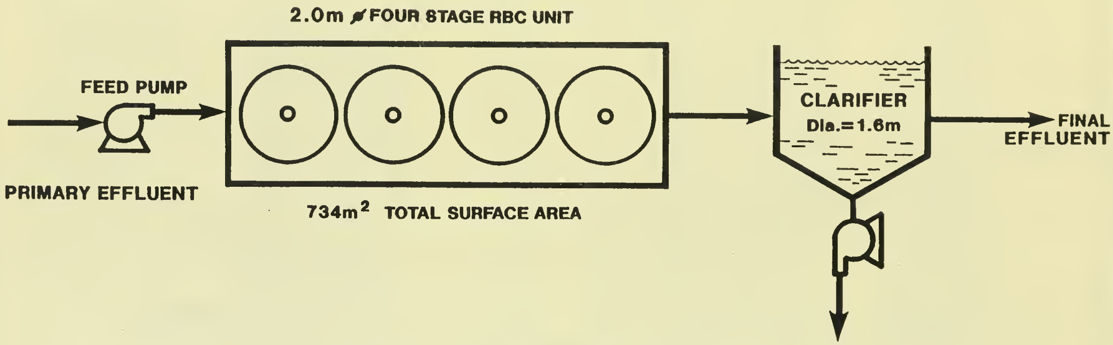


FIGURE 2.4: RBC PILOT PLANT

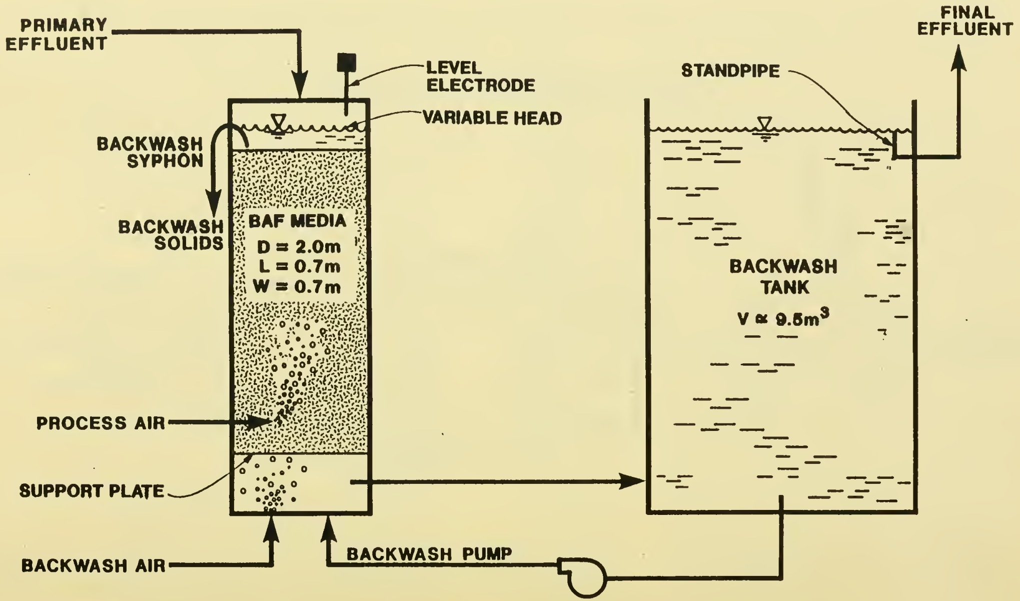


FIGURE 2.5: BAF PILOT PLANT

ASP PROCESS

To investigate the ASP process, a bench scale unit was used. The unit was used to confirm the results obtained from the City's Little River PCP and to compare with the results from the other pilot plants. Table 2.5 indicates the initial loading parameters used for the ASP pilot plant. In addition to operating the pilot plant at conditions similar to Plant No. 1 at the Little River PCP, an anoxic zone was added upstream of the aeration tank.

An anoxic environment is generally defined as an environment where molecular oxygen is absent but oxidized nitrogen (nitrite, nitrate) is present. The anoxic zone provides a short contact (0.33 h) time and initial high F/M ratio. Nitrate is supplied by both the return sludge and a nitrate-rich mixed liquor recycle from the aeration basin. Within the anoxic zone, floc-formers may use nitrate as an electron acceptor in the absence of molecular oxygen. This allows these organisms to exist in an exogenous phase in the high F/M anoxic region. Since filamentous organisms are obligate aerobes, they will not become active until the subsequent aeration step. At the aeration step most of the readily degradable, soluble organics have already been removed from the wastewater. This helps to prevent the filamentous organism population from far exceeding a balanced limit and causing a sludge bulking problem. Conditions favour the formation of floc-formers in the system at the expense of filamentous organisms.

Feed for the ASP was from the main feed well via a variable speed peristaltic pump. Flow through the anoxic zone, aeration tank and final clarifier was by gravity. RAS was returned to the anoxic zone, along with the MLSS. Wasting was accomplished batchwise from the aeration tank. The equipment sizes and requirements are shown in Figure 2-6.

SCHEDULE

A chronological listing of key milestone dates for the operation of the pilot plants and the study are summarized in Appendix A. As discussed earlier, the pilot plants were put into operation beginning in late June 1990; the last pilot plant was operational by August 1990. The pilot plant units were all operated continuously until the end of May 1991. The TF/SC and BAF pilot plants continued to be operated by the City to investigate additional operating conditions until the beginning of August 1991.

Each of the pilot plants treated a diurnal flow rate, modelled after the full-scale plant influent flow. The simplified flow pattern was based on visual inspection of the influent flow to the full-scale plant on a number of typical days. In general, the flow rate increases around 8 a.m. and, after a steep increase, levels off and is maintained until the late afternoon. The peak daily flow rate was estimated at twice the evening and early morning flow and 50 percent higher than the daily average flow. The peak flow was estimated to occur for one-third of the day from 8 a.m. to 4 p.m. Figure 2-7 shows

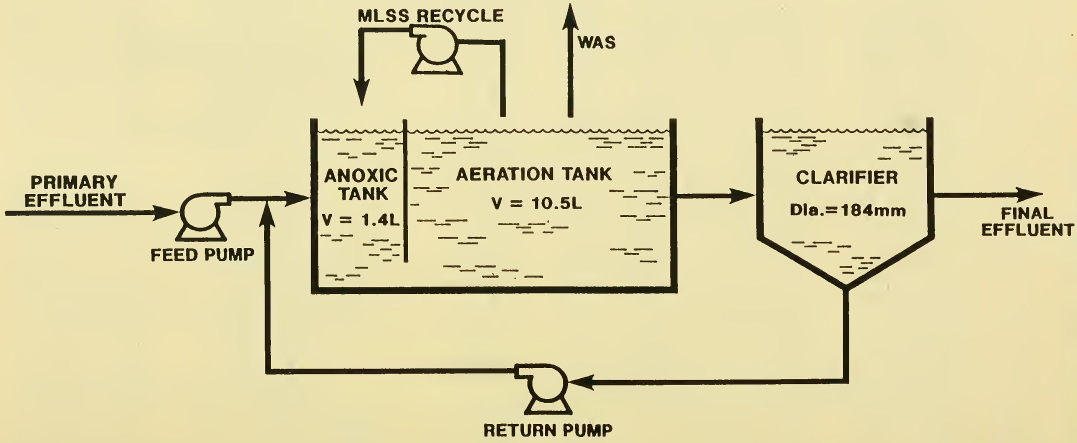


FIGURE 2.6: BENCH SCALE ACTIVATED SLUDGE PLANT

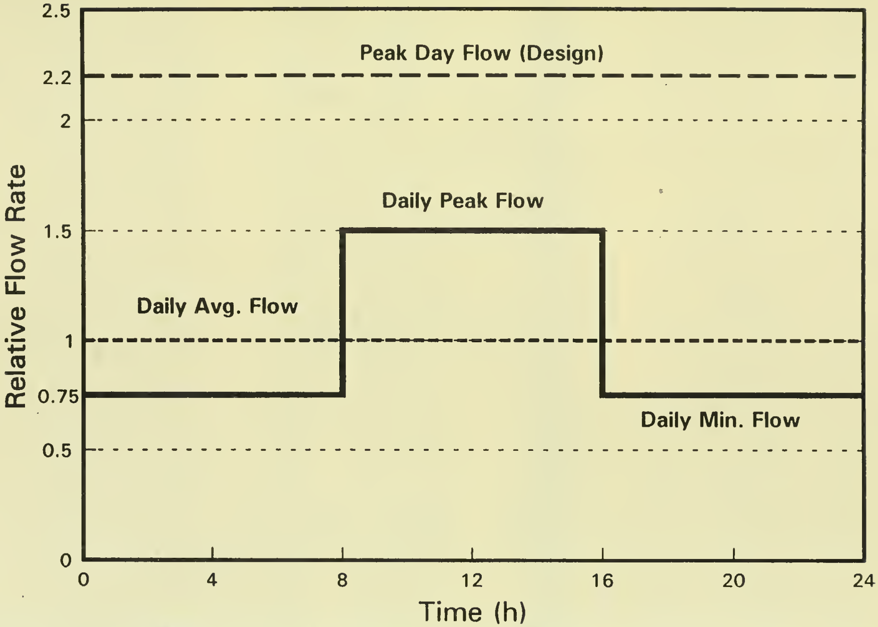


Figure 2-7 SIMPLIFIED DIURNAL FLOW PATTERN USED AS INFLUENT TO THE PILOT PLANTS

the simplified flow pattern used five days a week for each pilot plant. Weekends were generally operated at the weekday lower flow rate. To investigate the ability of each process to handle peak day flows, each pilot plant was operated at the full-scale plant's design peak factor. The full-scale plant is designed to treat a 2.2 peak factor, or 2.2 times the average daily flow. On days when the full-scale plant was treating peak flows, flow to the pilot plants was increased to 2.2 times the average flow, when possible.

Each pilot plant was initially operated at the process loadings that were determined as part of the review of each process technology. A number of changes were initiated, as a result of the operation of each pilot plant, to optimize the performance of each unit. These changes led to periods during which each process was operating at what was determined to be the "optimal" condition. The periods when each unit was operating as designed and optimized are presented in Table 2.6.. The operating periods for each unit are briefly discussed below.

Table 2.6
Optimization Periods for Operation of Each Pilot Process

Pilot Process	Periods of Optimization
TF/SC	• December 18, 1990 - May 1, 1991 • June 1, 1991 - August 2, 1991 • September 17, 1991 - April 26, 1992
RBC	• July 5, 1990 - May 31, 1991
BAF	• August 20, 1990 - August 2, 1991 (excluding period January 11, 1991 - March 11, 1991) • January 15, 1992 - April 26, 1992
ASP	• November 16, 1990 - May 31, 1991 (excluding the period December 22, 1990 to January 11, 1991)

TF/SC PROCESS

The TF began operation on July 6, 1990 with the effluent discharged directly to drain. The solids contactor (SC) and clarifier construction was finished on July 19, 1990. The SC was seeded on three occasions with activated sludge from the Little River PCP. After two solids losses from the SC, the TF was operated without the SC. The unit operated in this mode (TF with final clarifier) for a two-month period. In November 1990, the SC was restarted and began operation. The clarifier effluent quality was inadequate; therefore, on December 18 a portion of the TF effluent was diverted to drain to reduce the hydraulic and solids loading on the final clarifier. In addition, one of the two SC units was taken out-of-service to maintain a constant hydraulic retention time. On January 23, 1991, the loading to the final clarifier was further reduced to enable the unit to treat peak day flows. Solids were lost from the SC on May 1, 1991, potentially due to improved efficiency of the TF although long-term SC deficiencies are further discussed later in this report. The SC was restarted on May 14, 1991 by increasing the TF loading in an attempt to increase the organic and solids loading to the SC. The unit continued operating until August 6, 1991 when it was initially shut down.

The TF/SC process was re-started in September 1992 to investigate the operation of the SC unit. During this period the loading to the TF tower was increased and the SC hydraulic retention time was reduced in an attempt to try and improve the performance of the overall unit. On January 28, 1992, modifications were made to operate the TF/SC in a Mode III configuration (i.e. with a solids reaeration tank). This configuration was determined to be unsuitable. The TF/SC pilot plant was operated until April 26, 1992.

RBC PROCESS

The RBC was started on June 20, 1990 with a feed rate of 45 L/min and a disc rotation speed of 1 rpm. The unit ran without interruption from this time until the unit was shut down on June 14, 1991. The RBC was obtaining 50 and 60 percent SS and total BOD_5 removal 6 days after startup. After 15 days, the RBC was obtaining 95 percent NH_3-N removal.

BAF PROCESS

The BAF unit was started on July 27, 1991 at a feed rate of 38 L/min. After a week of operation, the viewing window for observing the media cracked and the unit had to be shutdown. The viewing window was repaired and the unit was restarted on August 15, 1991. The BAF was obtaining 50, 88, and 75 percent removal of SS, total BOD_5, and TP, as well as 43 percent NH_3-N removal, 16 days after startup.

Due in part to potential hydraulic limitations through the filter, the supplier initiated a number of vigorous backwashes around October 30 and again around December 10, 1990. As a result of these backwashes and loss of biomass inventory, nitrification through the filter was interrupted. A number of efforts were initiated to re-establish nitrification in the unit from January 11 to March 11, 1991 when wastewater temperatures were low. These attempts were unsuccessful. Partial nitrification was evident beginning in early April 1991. The supplier discovered that the internal air header for the unit's process air was broken in July 1991 and this air header was fixed. It was unknown when this line had broken during the study. The unit was operated until August 2, 1991. During July 1991, attempts were made to chill the influent wastewater to winter conditions to verify cold temperature nitrification rates. Unfortunately, the chilling unit was not capable of reducing the temperature to the levels required.

The BAF was re-started in September 1991 to investigate the ability of the unit to nitrify consistently and handle peak day flows. On January 15, 1992 the media depth was increased to 2.5 m, to assist nitrification in the unit. Peak hydraulic tests (i.e. 3 to

4 day duration) were performed in March and April 1992. The BAF pilot was operated until April 26, 1992.

ASP PROCESS

The unit was started up on July 6, 1991. The anoxic tank located upstream of the aeration tank was added on September 13, 1991. Due to poor settling quality of the effluent from the unit, the aeration section HRT was increased to 3.5 h and a rake device was added to the final clarifier on November 16, 1991. The unit's feed lines froze on December 24, 1991, due to cold ambient temperatures. The unit was restarted on January 11, 1991 with provision for temperature control. The unit was shut down on June 14, 1991.

The ASP pilot plant was obtaining essentially complete nitrification after a month of operation. Due to the problems discussed above, the unit was obtaining poor SS, total BOD_5, and TP removal until November, 1991.

PILOT PLANT ANALYTICAL SAMPLING AND MONITORING

Pilot plant performance was monitored closely on a daily basis. Twenty-four hour time-proportioned automatic samplers were located on the pilot plant influent, each pilot plant effluent, and the effluent of the trickling filter tower. In general, daily composite samples were analyzed for SS, TP, and BOD_5. The nitrogen series was analyzed every second day. Additional samples were taken on the trickling filter effluent and the pilot plant influent. Table 2.7 outlines the sampling program used.

In addition to daily composite sampling, ammonia profiles through the RBC, TF/SC and BAF were done on a periodic basis. Daily monitoring was conducted to observe the performance of the pilot plants. A complete daily monitoring sheet is provided in Appendix C. In general, influent and effluent wastewater temperatures, effluent dissolved oxygen concentration, and flow rates were monitored daily (five times/week) at all the pilot plants.

To ensure the accuracy of the analytical sampling, a quality assurance/quality control (QA/QC) program with CANVIRO Analytical Laboratories (CALL) was conducted twice each month. In addition, a three-way analytical check was conducted with the City of Windsor, MOE, and CALL laboratories on three occasions. The results of this QA/QC program are summarized in Table 2.8. The QA/QC program between the City of Windsor and CALL was compared by conducting a paired t-test[16]. The results of this test indicated that the two series of sample analyses were the same based on a 95 percent confidence interval.

Table 2.7
Analytical Sampling Schedule

Parameter	Frequency	Sample Location
pH	Daily	• Influent, TF effluent, effluent of TF/SC, RBC, BAF and ASP
SS	Daily	• Influent, TF effluent, effluent of TF/SC, RBC, BAF and ASP
Total BOD_5	Daily	• Influent, effluent of TF/SC, RBC, BAF and ASP
Soluble BOD_5	Daily until end of January 1990, then every third day	• Influent, TF effluent
Carbonaceous BOD_5	Daily until end of January 1990, then every third day	• Effluent of TF/SC, RBC, BAF and ASP
Nitrogen Series (NH_3-N, TKN & NO_3 + NO_2-N)	Every two days	• Influent and effluent of TF/SC, RBC, BAF and ASP
NH_3-N	Daily	• TF effluent

Table 2.8
QA/QC Results During Pilot Plant Testing Program

Date	Location	Concentration (mg/L)																	
		TSS			TBOD5			TP			TKN-N			NH3-N			NO3-N		
		CITY	CALL	MOE	CITY	CALL	MOE	CITY	CALL	MOE	CITY	CALL	MOE	CITY	CALL	MOE	CITY	CALL	MOE
24-Aug-90	PRIM EFF	23	17	NA	37	41	NA	0.8	0.8	NA	12.2	14.2	NA	8.5	9.8	NA	NA	NA	NA
07-Sep-90	TF EFF	28	31	NA	12	12	NA	0.6	0.8	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA
14-Sep-90	RBC EFF	4	5	NA	6	4	NA	0.0	0.2	NA	2.8	2.4	NA	1.2	1.0	NA	4.0	4.3	NA
25-Sep-90	BAF EFF	3	3	NA	6	<6	NA	0.1	0.1	NA	2.3	1.4	NA	1.2	0.4	NA	8.7	5.6	NA
14-Oct-90	TF FINAL EFF	10	20	NA	6	9	NA	0.4	0.4	NA	4.0	2.0	NA	1.6	0.4	NA	9.2	NA	NA
07-Nov-90	PRIM EFF	27	26	NA	46	42	NA	1.0	0.9	NA	13.1	9.7	NA	6.0	6.2	NA	0.2	0.2	NA
12-Nov-90	PRIM EFF	36	27	NA	44	21	NA	1.0	1.4	NA	14.9	10.3	NA	9.7	5.8	NA	0.2	0.2	NA
12-Nov-90	RBC EFF	7	15	NA	7	10	NA	0.5	0.5	NA	2.8	1.7	NA	0.4	1.8	NA	9.6	5.6	NA
12-Nov-90	TF/SC EFF	21	28	NA	16	17	NA	0.5	1.2	NA	4.8	2.5	NA	0.6	2.3	NA	9.3	5.9	NA
09-Jan-91	RBC EFF	8	9	NA	14	5	NA	0.3	0.2	NA	5.6	5.2	NA	3.3	3.7	NA	6.5	4.1	NA
09-Jan-91	BAF EFF	5	7	NA	7	<6	NA	0.3	0.2	NA	11.1	9.7	NA	9.2	9.1	NA	2.3	0.7	NA
15-Jan-91	TF EFF	18	24	NA	8	6	NA	NA	0.5	NA	NA	4.0	NA	2.0	2.3	NA	NA	NA	NA
15-Jan-91	TF/SC EFF	12	17	NA	10	4	NA	0.5	0.5	NA	2.8	2.2	NA	0.5	<.02	NA	9.2	6.4	NA
05-Feb-91	PRIM EFF	32	34	27	113	68	57	0.6	0.8	0.7	16.3	15.4	15.2	8.9	9.1	9.3	0.2	0.2	<0.1
05-Feb-91	RBC EFF	17	20	20	15	15	17	0.4	0.7	0.4	5.6	4.9	5.8	2.6	2.6	2.6	6.3	4.9	6.4
27-Feb-91	RBC EFF	6	10	9 *	3	5	10 *	0.2	NA	0.3 *	3.3	3.1	3.0 *	0.6	1.6	1.3 *	12.8	8.3	8.6 *
17-Mar-91	PRIM EFF	30	18	NA	33	18	NA	0.7	0.9	NA	20.9	14.7	NA	12.6	12.4	NA	0.6	<0.9	NA
17-Mar-91	RBC EFF	11	9	NA	9	5	NA	0.3	0.4	NA	4.5	1.5	NA	1.5	0.6	NA	14.0	12.0	NA
17-Apr-91	PRIM EFF	21	15	NA	47	35	NA	0.7	0.4	NA	15.5	16.2	NA	9.0	11.0	NA	0.4	<0.2	NA
17-Apr-91	RBC EFF	6	4	NA	9	4	NA	0.2	0.2	NA	5.1	2.6	NA	1.9	1.8	NA	1.9	7.8	NA
23-Apr-91	RAW INFL	NA	NA	109	NA	NA	66	NA	NA	2.6	NA	NA	14.3	NA	NA	8.8	NA	NA	<0.1
23-Apr-91	PRIM EFF	24	NA	43	46	NA	64	0.6	NA	1.4	11.6	NA	12.8	7.6	NA	8.6	0.5	NA	0.8
23-Apr-91	FINAL EFF	NA	NA	8	NA	NA	5	NA	NA	0.1	NA	NA	0.7	NA	NA	<0.1	NA	NA	8.7
09-May-91	PRIM EFF	23	25	NA	58	49	NA	0.5	0.4	NA	14.5	13.3	NA	10.5	11.3	NA	0.3	<0.2	NA
09-May-91	RBC EFF	8	9	NA	13	<20	NA	0.2	0.2	NA	6.3	3.4	NA	3.1	2.5	NA	7.4	5.8	NA
21-May-91	PRIM EFF	31	NA	NA	77	76	NA	0.7	NA	NA	18.1	NA	NA	11.1	NA	NA	0.8	NA	NA
27-May-91	PRIM EFF	30	30	38 *	50	59	54 *	0.9	0.8	1.0 *	13.2	12.7	13.6 *	7.9	9.0	8.2 *	0.5	<0.2	0.5
27-May-91	RBC EFF	12	9	13 *	13	6	15 *	0.3	0.3	0.4 *	4.6	3.9	4.2 *	2.0	1.7	1.6 *	6.0	4.3	6.2 *

Notes :
CITY - City of Windsor Lab
CALL - CANVIRO Analytical Laboratories Ltd.
MOE - Ontario Ministry of the Environment Lab
NA refers to not applicable
* - Indicates average of sample and duplicate results

Section 3
RESULTS OF PILOT PLANT OPERATIONS

The average monthly performance of the liquid train full-scale plant for the study period is summarized in Table 3.1. The full-scale plant treated an average daily flow of 129,000 m^3/d or 81 percent of the plant's rated hydraulic capacity. The full-scale plant met its effluent compliance requirements for SS, and total BOD_5, based on yearly removal and TP on a monthly basis during this period. The effluent SS, total BOD_5, and TP concentrations averaged 23, 36 and 0.6 mg/L.

Table 3.1
Full-Scale Plant Performance During Pilot Testing

Month	Treated Flow (1000 m^3/d)	Monthly Average Concentrations					
		Influent			Effluent		
		TSS	$TBOD_5$	TP	TSS	$TBOD_5$	TP
Jun 90	126.3	179	118	4.2	13	48	0.5
Jul 90	111.9	172	85	4.6	14	26	0.4
Aug 90	129.9	147	86	3.6	16	25	0.5
Sep 90	162.8	133	81	3.5	16	24	0.6
Oct 90	155.7	132	97	3.7	19	32	0.7
Nov 90	126.6	157	111	4.5	25	36	0.8
Dec 90	162.6	148	121	4.0	34	39	0.8
Jan 91	127.9	140	112	3.8	23	39	0.6
Feb 91	117.5	125	117	3.6	25	38	0.6
Mar 91	115.9	143	115	3.8	25	36	0.6
Apr 91	138.1	131	96	3.4	25	38	0.6
May 91	128.7	153	116	3.4	35	51	0.8
Jun 91	130.8	92	92	3.3	25	31	0.5
Jul 91	110.2	137	110	4.5	25	41	0.7
Aug 91	119.0	113	89	3.2	24	28	0.6
Sep 91	105.0	143	99	4.1	25	37	0.7
Oct 91	118.5	165	148	3.8	30	46	0.8
Nov 91	113.4	146	137	4.1	27	36	0.6
Dec 91	111.7	168	126	3.6	21	35	0.4
Jan 92	115.8	203	166	4.4	19	41	0.4
Feb 92	143.0	145	124	3.3	23	37	0.5
Mar 92	144.8	153	114	3.6	20	31	0.5
Apr 92	151.6	148	108	3.5	18	28	0.4
Average	129.0	147	112	3.8	23	36	0.6

During the testing, the pilot plants treated a portion of the effluent from Clarifiers No. 5 and 6 at the full-scale plant. Table 3.2 compares the difference between the full-scale plant final effluent (i.e. entire plant final effluent) and the pilot plant influent concentrations for SS, total BOD_5, and TP from August 1990 to April 1992 (excluding August 1991 when the processes were out-of-service). The results show that on average the SS from the full-scale plant were similar to those sampled from the influent to the pilot plants. The total BOD_5 and TP were on average about 20 percent higher for the pilot influent than the effluent from the full-scale plant. From the end of October 1990 to mid-January 1991, polymer addition to full-scale Clarifiers No. 5 and 6 was discontinued to increase the trickling filter loading in an attempt to startup the SC unit and to determine the impact of increased loadings on the other pilot plants. As a result, the pilot plant influent concentrations increased for November, December, and January. The results indicated that the influent to the pilot plants are representative of the effluent from the full-scale plant.

The remainder of this section focuses on the performance of the pilot plants over the study period. The section is divided into six parts:

- pilot plant effluent requirements
- a review of the characteristics of the pilot plant influent, including organic and nitrogen concentrations and loadings, pH, and temperature
- a discussion of each pilot plant's performance during the testing period, highlighting the effluent parameters and relationships common and unique to each process
- a bench-top review of waste solids settling and dewaterability testing of the waste solids
- a discussion of observed operational and maintenance needs for each process
- a summary of the performance of each pilot plant, with emphasis on comparing their performance

EFFLUENT REQUIREMENTS

Input from the MOE was sought to determine the possible effluent requirements for the upgraded facility. The MOE indicated to the City that, at the present time, the expected effluent design requirements for the upgraded plant would not be more stringent than the following:

- 15 mg/L SS
- 15 mg/L total BOD_5
- 0.5 mg/L TP
- 0.1 mg/L un-ionized ammonia-N.

Table 3.2
Comparison of Full-Scale Plant Effluent and Pilot Plant Influent

Month	Monthly Average Concentrations (mg/L)					
	Full-Scale Plant Effluent			Pilot Plant Influent		
	TSS	TBOD5	TP	TSS	TBOD5	TP
Aug-90	16	25	0.5	14	29	0.6
Sep-90	16	24	0.6	11	28	0.6
Oct-90	19	32	0.7	16	35	0.7
Nov-90	25	36	0.8	29	43	1.2
Dec-90	34	39	0.8	32	51	1.2
Jan-91	23	39	0.6	21	44	0.7
Feb-91	25	38	0.6	20	53	0.6
Mar-91	25	36	0.6	22	39	0.7
Apr-91	25	38	0.6	24	44	0.7
May-91	35	51	0.8	29	49	0.8
Jun-91	25	31	0.5	26	41	0.6
Jul-91	25	41	0.7	24	46	1.0
Aug-91	24	28	0.6	NA	NA	NA
Sep-91	25	37	0.7	25	49	0.7
Oct-91	30	46	0.8	26	60	0.8
Nov-91	27	36	0.6	28	55	0.9
Dec-91	21	35	0.4	22	50	0.4
Jan-92	19	41	0.4	25	60	0.6
Feb-92	23	37	0.5	22	42	0.5
Mar-92	20	31	0.5	23	36	0.6
Apr-92	18	28	0.4	20	30	0.5
Average*	24	36	0.6	23	44	0.7

notes : 1) * August 1991 not included in average, since pilot plants were out-of-service

The discussion between the MOE and the City is provided in the meeting notes of October 30, 1990 (Appendix A).

The un-ionized NH_3-N effluent requirement would be based on the end-of-pipe discharge conditions in terms of wastewater temperature and pH. After project initiation, the Steering Committee recommended a target effluent total NH_3-N concentration of 2 mg/L be used to examine actual nitrification performance. During the extended period of operation (September 17, 1991 to April 26, 1992) the effluent ammonia-N target was reassessed at 3 mg/L.

During pilot testing, adjustments were made to processes that were unable to meet the effluent design objectives on a consistent basis. The target effluent total NH_3-N concentration was monitored; however, process changes were not made if units did not meet this target consistently.

PILOT PLANT INFLUENT CHARACTERISTICS

The monthly average influent characteristics are shown in Table 3.3. Raw data for the pilot plant influent characteristics and monitoring data are contained in Appendix E. Probability plots of each pilot plant's effluent objective parameters are contained in Appendix F. Influent conditions were divided into five periods, based on the operating conditions of the two primary clarifiers feeding the pilot plants and the extension period. These periods were:

Period 1	July 2/90 to October 29/90	alum and polymer used in full-scale plant
Period 2	October 30/90 to January 10/91	alum only, and clarifier No. 5 taken out-of-service
Period 3	January 11/91 to May 31/91	ferric and reduced polymer
Period 4	June 1/91 to August 2/91	ferric only
Period 5	September 17/91 to April 26/92	ferric and reduced polymer addition

Table 3.3
Average Pilot Plant Influent Characteristics

Period	Temperature (°C)	pH	Concentration (mg/L)						
			SS	Total BOD_5	Soluble BOD_5	TP	TKN	NH_3-N	NO_3-N
Overall Period (Jul. 2/90 to Apr. 26/92)	16	7.1	23	43	39	0.9	14.0	9.4	0.4
Alum + Polymer Addition (Jul. 2/90 to Oct. 29/90)	22	7.1	13	31	32	0.6	12.0	8.0	0.3
Alum only (Oct. 30/90 to Jan. 10/91)	15	7.3	31	46	33	1.2	15.0	9.2	0.6
Ferric and Polymer (Jan 11/91 to May 31/91)	13	7.1	23	46	39	0.7	15.2	9.6	0.5
Ferric Only Jun. 1/91 to Aug. 2/91	23	7.0	25	43	38	1.6	13.8	9.6	0.3
Ferric and Polymer Sep. 17/91 to Apr. 26/92	14	7.0	24	47	49	0.6	14.5	9.9	0.4

The normal operation of the two primary clarifiers feeding the pilot plants was changed on October 30, 1991, to increase the loading to the pilot plants. This change was initiated to assist the TF/SC process's solids contactor startup and to determine the impact on the other pilot plants. One of the two primary clarifiers was also taken out-of-service to assist in increasing the load to the pilot plants without adversely affecting the plant's effluent quality. The full-scale plant changed from alum to ferric in January 1991 as part of the plant's yearly chemical contract. Polymer addition to the two clarifiers feeding the pilot plants was reduced during Period 3 to maintain a higher loading especially to the TF/SC process. The impact of no polymer addition was evident in the pilot plant influent data when compared with the full-scale plant effluent data. Pilot plant influent concentrations for Period 2 were 35 to 33 percent higher than the overall average for SS and TP (Table 3.3). The total BOD_5 concentration was only 7 percent greater during this period as compared to the overall period, most likely because the total BOD_5 concentration during Period 3 remained unchanged with the re-introduction of polymer. The total BOD_5 concentration during Period 2 was about 50 percent greater than the concentration during Period 1. The total BOD_5 concentration during Period 3 appears to have remained unchanged from Period 2, due to an increase in the soluble BOD_5 concentration. The soluble BOD_5 concentration during Period 2 averaged 33 mg/L, whereas during Period 3 the concentration increased to 39 mg/L.

The pilot plant influent NH_3-N concentration averaged 9.4 mg/L for the 23-month period. The concentration varied from 2 to 30 mg/L; however, the pilot plant influent NH_3-N concentration was greater than 16.8 mg/L or less than 5.1 mg/L only 10 percent of the time . High influent NH_3-N concentrations did occur around February 17, 1991. NH_3-N concentrations greater that 15 mg/L were believed to have occurred due to a rapid in-flush of urea into the collection system. Urea was used for ice control in the downtown area and, during this period, a rainfall caused urea input (as ammonia) to be washed down to the plant through the combined collection system. The West Windsor PCP currently does not have an effluent NH_3-N criteria, and if and when the plant has this criteria, the City has indicated that it will re-examine the use of urea for ice control. The nitrate concentration averaged 0.4 mg/L, but ranged between 0.1 and 3.1 mg/L over the study.

The pH of the pilot plant influent varied from 6.6 to 7.7 over the testing period, averaging 7.1 on a H^+ concentration basis. pH analyses were conducted using the daily composite samples. Since pH can vary with time and in the sample bottle, a check was conducted using an online calibrated pH meter. This meter was located in the main feed well, the RBC effluent and TF effluent for a total of five days. The results, summarized in Table 3.4, show that the average daily pH determined by the discrete online reading every four hours compared closely with the pH determined from the composite samples for the same day. In addition, the discrete online pH measurements showed that the pH varied by 0.1 to 0.5 units throughout the day in the process effluents.

Table 3.4
Comparison of Daily Composite and Online pH Measurement

Date 1991	Sample Location	Average* Online	Online Min - Max	Plant Lab**
May 21	RBC Effluent	7.1	6.9 - 7.4	7.4
May 22	PE 5 and 6	6.9	6.8 - 6.9	7.1
May 23	TF Effluent	7.3	7.0 - 7.5	7.3
May 24	TF Effluent	7.3	7.2 - 7.4	7.1
May 25	TF Effluent	7.2	7.0 - 7.5	7.1
May 26	TF Effluent	7.3	7.0 - 7.4	7.2

Notes: * based on average of 6 discrete readings (every 4h)
** 24-h composite sample

The pilot plant influent wastewater temperature was taken five times a week. Figure 3-1 shows a trend plot of the wastewater temperature readings to April 26, 1992. The monthly average wastewater temperatures are shown in Table 3.5. The temperature started at approximately 24°C in the fall of 1990, and by mid-December the wastewater temperature was averaging 13°C. The minimum wastewater temperature of 10 to 11°C occurred from about mid-January to mid-March. The wastewater temperature began rising in March and by the end of May 1991 was averaging 18°C. The temperature began to lower again in November 1991 and averaged 11°C for January and February 1992. The wastewater temperature was 13 to 14°C at the end of the study in April 1992.

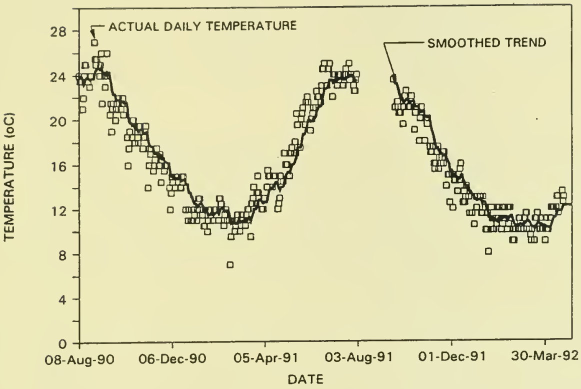


FIGURE 3.1 PILOT PLANT INFLUENT WASTEWATER TEMPERATURE

Table 3.5
Monthly Average Pilot Plant Influent Wastewater Temperatures

Month		Average Temperature (°C)
1990	August	24
	September	22
	October	19
	November	16
	December	13
1991	January	11
	February	11
	March	11
	April	14
	May	18
	June	22
	July	24
	August	--
	September	21
	October	21
	November	16
	December	14
1992	January	11
	February	11
	March	10
	April	12

Notes:
1. -- indicates no values recorded

PERFORMANCE OF PILOT PLANTS

The performance of each pilot plant over the evaluation period is detailed below. The performance of each pilot plant in terms of its effluent performance and hydraulic loading is discussed and relationships between process loading rates and effluent concentration are investigated. Raw analytical data and pilot plant monitoring data for each process are contained in Appendix E. Appendix F contains probability plots of the effluent objective parameters for each process.

TF/SC PROCESS

General

The operating periods of the TF and the TF/SC process were outlined earlier in this report. This section focuses on the periods when the process was operating as a coupled TF/SC. These periods were:

- December 18, 1990 to May 1, 1991
- June 1, 1991 to August 2, 1991
- September 17, 1991 to April 26, 1992

The period from August 30 to November 4, 1990, when the process was operated as a TF with a final clarifier, is also discussed. Table 3.6 presents the average influent parameters (i.e. wastewater flows, pH, and temperature) and effluent concentrations for these four periods and the overall period. The three periods when the unit was operating as a TF/SC are presented separately for comparison, since the hydraulic loading rate to the TF was significantly higher during these periods. An additional period from November 13, 1991 to January 24, 1992, when the TF/SC and TF with a final clarifier were operated in parallel, will also be discussed.

Effluent Quality

Table 3.6 indicates that during all three periods the TF and the TF/SC met the effluent objectives for TP and un-ionized NH_3-N on an average basis. The target process total NH_3-N concentration of 2 mg/L was also met except during the follow-up extension period (September 17, 1992 to April 16, 1992). The TP concentration exceeded the 0.5 mg/L objective during the initial extension period due to a high influent TP concentration of 9 mg/L that occurred in the last week of July. If this anomaly is removed, the average TP concentration for this period was 0.4 mg/L.

Both the TF and the TF/SC were marginal in meeting the objectives for total BOD_5 and SS. The average effluent SS concentration averaged 14 to 15 mg/L for the TF and TF/SC processes during the first year of operation. The average total BOD_5 concentration exceeded 15 mg/L for three of the periods at 16, 18, and 26 mg/L. The average total BOD_5 concentration during the period from December 18, 1990 to May 1, 1991 was at the objective concentration of 15 mg/L. The carbonaceous BOD_5 concentration was less than 15 mg/L for all four periods; therefore, higher total BOD_5 concentrations were probably due to nitrogeneous oxygen demand in the sample bottle. Under the operational modes examined in this study, it is evident that the TF/SC provides no safety factor in meeting the design objectives for BOD_5, SS, and TP.

During April 1991, the wastewater temperature rose from 10°C to 16°C. During this period, the solids from the final clarifier and SC tank were lost. At the end of May 1991, the hydraulic loading to the TF was increased by 100 percent until the beginning of August 1991, in an attempt to increase the loading to the SC. Also, an anionic polymer was added to the SC effluent to assist in capturing the solids and starting the SC. The polymer was started on May 28, 1991 at a dosage of 0.4 mg/L and the dosage was reduced to about 0.2 mg/L by mid-June. The polymer was shut off on July 3, 1991. This event is presented in more detail later in the report (sub-section SC Unit Solids Loss, page 3-23). During the follow-up extension period (September 17, 1991 to April 26, 1992) the loadings to the TF tower were varied to assist the operation of the SC tank.

Table 3.6
Performance Summary of the TF/SC Pilot Plant

Period	Influent Parameters				Effluent Concentration (mg/L)					
	TF Flow (L/min) [TF Wetting Rate $m^3/m^2 \cdot d$]	SC Flow (L/min)	pH	Temp (°C)	SS	Total BOD_5	Carb. BOD_5	TP	NH_3-N	Un-ionized NH_3-N
Overall Period When Operated as TF/SC••	167 [40.2]	71	7.4	14	16	20	8	0.6• (0.5)	3.3	0.036
Trickling Filter Only (Aug. 30/90 to Nov. 4/90)	172 [41.7]	-	7.2	20	14	18	9	0.5	2.0	0.017
TF/SC Operation (Dec. 18/90 to May 1/91)	132 [32.0]	53	7.4	11	15	15	6	0.4	1.1	0.007
Initial Extension Period - TF/SC Operation (Jun. 1/91 to Aug. 2/91)	229 [55.5]	53	7.2	23	14	16	7	0.7• (0.4)	1.3	0.014
Follow Up Extension Period - TF/SC Operation (Sep. 17/91 to Apr. 26/92)	171 [41.5]	88	7.5	13	17	26	13	0.5	4.8•••	0.054

Notes:

• Elevated average TP concentration due to high influent concentration period for July 20 to 21, 1991 when concentrations were greater than 15 mg/L.
Value in brackets has elevated TP concentrations removed.

•• Overall period is average of three operating period, these are:
Dec. 18/90 to May 1/ 91 (TF/SC)
Jun. 1/91 to Aug. 2/91 (TF/SC)
Sep. 17/91 to Apr. 26/92 (TF/SC)

•••Optimizing SC operation, effluent quality deteriorated due to action initiated to optimize SC, complete nitrification was not an objective during this period.

Near the end of the initial extension period the hydraulic loading to both the TF tower and final clarifier were increased to an average of 67.0 and 32.0 $m^3/m^2 \bullet d$ respectively for an eight-day period. This test was initiated to determine the impact of an increased load to the TF tower and a decreased HRT in the SC unit. The process did not perform well during this eight-day period and exceeded the effluent design objective concentrations. Effluent concentrations averaged 18, 21, 0.7, and 2.8 mg/L for SS, total BOD_5, TP, and ammonia-N for this period.

Hydraulic Loading

On average the TF tower hydraulic loading was greater than the target loading of 29 $m^3/m^2 \bullet d$, averaging 41.7, 32.0, 55.5 and 41.5 $m^3/m^2 \bullet d$ for the four periods. On a peak day basis, the maximum daily flow rates were 61.7, 48.7, 79.9, 97.0 $m^3/m^2 \bullet d$ for the four periods. This corresponds to a 2.4 peak factor over the average flow treated, or a 3.3 peak factor above an average daily flow of 29.4 $m^3/m^2 \bullet d$. The hydraulic loading to the SC and final clarifier was reduced for the period when the SC was operating. During this period the average SC flow was 53 L/min corresponding to 38 min HRT with a peak daily flow of 95 L/min (from December 18, 1990 to August 2, 1991). Therefore, the SC and final clarifier was operated at a 1.8 peak factor. Average clarifier SOR was 16.9 $m^3/m^2 \bullet d$ with a peak day of 32.1 $m^3/m^2 \bullet d$ (from December 18, 1990 to August 2, 1991). During the follow-up extension, the SC and final clarifier were operated at peak flow continuously. This mode of operation was used to reduce the SC HRT to 25 min.

Both the TF tower and the SC (when operated) were able to treat a simplified diurnal flow pattern modelled on the flow treated by the full-scale plant.

The operating parameters for the four periods and the overall period are summarized in Table 3.7. The design TF loading rates, that is the total organic loading (TOL) and the TF wetting rate, exceeded the initial target values of 0.25 kg $BOD_5/m^3 \cdot d$ and 29.4 $m^3/m^2 \bullet d$, especially during the period when the TF was operated on its own (August 30 to November 1990), during the initial extension period of the TF/SC (June 24 to August 6, 1991) and during the follow-up extension period (Sept. 17/91 to Apr. 26/92). The wetting rate averaged 41.7, 55.5, and 41.5 $m^3/m^2 \bullet d$ during these three periods, respectively. The SC solids retention time (SRT) averaged about 2.2 d during the periods when the SC was operating. The SC hydraulic retention time averaged 38 minutes during the first two periods and 25 minutes during the last period.

TF Versus TF/SC

When comparing the periods when the pilot was operated as a TF with a final clarifier and the period when it was operated as a TF/SC, a number of observations are significant. The average effluent SS and TP concentrations for both modes of operation are similar, averaging around 14 to 15 mg/L for SS and 0.4 to 0.5 mg/L for TP. The results when only the TF was operated are significant because, during this period, the clarifier hydraulic loading was higher than the overall average. The final clarifier SOR was the highest, averaging 55 $m^3/m^2 \bullet d$ during TF operation. The TF wetting rate averaged 41.7 $m^3/m^2 \bullet d$.

Table 3.7
Summary of TF/SC Influent Loadings and Operating Parameters

Period	Trickling Filter			Solids Contactor		Final Clarifier
	Total Organic Loading (kg $BOD_5/m^3 \cdot d$)	Nitrogen Loading (kg $NH_3\text{-}N/m^3 \cdot d$)	Wetting Rate ($m^3/m^2 \cdot d$)	SRT (d)	HRT (min)	Surface Overflow Rate ($m^3/m^2 \cdot d$)
Overall Period (Jun. 26/90 to Apr. 26/91)	0.42	0.10	40.2	-	-	29.2
Trickling Filter Only (Aug. 30/90 to Nov. 4/90)	0.32	0.08	41.7	-	-	55.0
TF/SC Operation (Dec. 18/90 to May 1/90)	0.33	0.07	32.0	2.4	38	16.9
Extension Period - TF/SC (Jun. 1/91 to Aug. 2/91	0.56	0.13	55.5	2.5	38	16.9
Follow Up Extension Period - TF/SC Operation (Sept. 17/91 to Apr. 26/92)	0.47	0.10	41.5	2.0	25	28.0

The effluent total BOD_5 and NH_3-N concentrations were lower when the SC was in operation. The effluent total BOD_5 concentration averaged 15 mg/L during the periods when the SC was operating, compared to 18 mg/L when only the TF was operating. This comparison is also reflected in the effluent carbonaceous BOD_5 concentration which averaged 7 mg/L when the SC was in operation, compared to 9 mg/L when only the TF was operating.

The effluent total NH_3-N concentration was also lower during the period when the SC was in operation. The effluent NH_3-N concentration averaged 1.1 mg/L when the SC was in operation, as opposed to 2.0 mg/L when only the TF was operating. The TF effluent total NH_3-N concentration averaged between 2.4 and 3.4 mg/L when the SC was in operation, indicating that a further 1.3 to 2.3 mg/L NH_3-N reduction occurred in the SC tank.

It is important to note that the period when the TF was operated on its own was in the fall of 1990 at higher wastewater temperatures. The average wastewater temperature during the period when only the TF was in operation was 20°C. Hydraulics would have to be decreased at lower wastewater temperatures during winter if the TF process alone was to maintain adequate nitrification. During the main period when the TF/SC was operating, from December 18, 1990 to May 1, 1991, the average wastewater temperature was 11°C.

A side-by-side comparison of the TF/SC and TF process was conducted during part of the follow-up extension period from November 13, 1992 through to January 24, 1992. During this period a portion of the TF tower effluent was directed to the SC and final clarifier and another portion to a separate final clarifier. The performance of each process over this 2-month period is summarized in Table 3.8. The TF tower loadings during this period were 0.52 kg $BOD_5/m^3{\bullet}d$ and 41.7 $m^3/m^2{\bullet}d$ for the TOL and wetting rate respectively. The clarifier SORs for each process were 29.1 and 23.5 $m^3/m^2{\bullet}d$ for the TF/SC and TF process respectively.

Table 3.8
Comparison of Effluent Performance for TF/SC and TF Processes
(November 13, 1991 to January 24, 1992)

Effluent Parameter	Average Concentration (mg/L)	
	TF/SC	TF
SS	17	13
$TBOD_5$	29	27
$CBOD_5$	14	16
TP	0.3	0.3
NH_3-N	5.3	7.2

Table 3.8 indicates that both processes were marginal in terms of meeting the SS and BOD_5 effluent objective. Overall, the TF/SC process did not perform any better than the TF process in terms of effluent SS, $TBOD_5$, $CBOD_5$ or TP. The TF/SC process did perform better in terms of ammonia-N removal due to an additional 2 mg/L ammonia-N removal in the SC tank. However, the main purpose of including the SC portion of the TF/SC process was to improve the final effluent SS concentration by improving settleability. Figure 3-2 is a trend plot comparing the effluent SS from both the TF/SC and the TF process. The smoothed SS trend plot indicates the effluent SS from each process is marginal with respect to the 15 mg/L objective and that the TF process effluent SS was generally better than the TF/SC process for this period.

Trends

Trend plots show the operation of the TF/SC from July 2, 1990 to April 26, 1992. Figure 3-3 is a trend plot of the flow rate to the TF tower and the SC unit. Continuous flow to the SC did not begin until November 5, 1991 when the SC was seeded with MLSS from the Little River PCP. From November 5 to December 17, 1990, the flow through the SC was the same as the flow through the TF tower. On December 18, 1990, part of the flow through the TF was diverted to reduce the SOR on the final clarifier to 30 $m^3/m^2 \cdot d$. It appeared that this flow rate was the maximum the clarifier could handle to achieve the required effluent SS concentration with a flocculant suspended growth-type suspension. Therefore, the average flow through the SC and clarifier was reduced on January 11, 1991 to 15 $m^3/m^2 \cdot d$. This reduction was implemented to provide the system with the ability to handle a 2.2 peak factor. During the follow-up extension the flow to the SC and final clarifier was increased to the maximum SOR to reduce the SC unit's HRT.

Figure 3-4 is a smoothed trend plot of the effluent SS from the TF and the final effluent SS of the pilot plant. Smooth data is based on a 7-d effective value. The data shown from this pilot plant were either settled TF effluent or settled SC effluent. The SS reduction between the TF tower effluent and final clarifier effluent increased when the SC unit was in service. For the period when the TF was operated on its own (i.e. with a final clarifier), a 36 percent SS reduction occurred between the TF tower effluent and final clarification. The TF tower effluent averaged 22 mg/L and the final effluent averaged 14 mg/L when the TF was operated on its own. When the SC was operating, the SS reduction increased to 46 percent. During this period, the TF tower effluent SS averaged 30 mg/L and the final effluent averaged 16 mg/L.

Figure 3-5 is a trend plot of the trickling filter effluent soluble BOD_5 concentration. The TF effluent soluble BOD_5 concentration averaged 12 mg/L. The trickling filter effluent soluble BOD_5 was less than 15 mg/L for 81 percent of the time. The primary effluent soluble BOD_5 concentration averaged 39 mg/L over the study period; therefore, the TF tower was obtaining a 70 percent removal efficiency of soluble BOD_5. The final effluent total BOD_5 concentration is shown in Figure 3-6. The effluent total BOD_5 concentration varied greatly; however, approximately 89 percent of the time the concentration was less than 25 mg/L during the first year of operations (July 2, 1990 to May 31, 1991). During the second year of operation the effluent total BOD_5

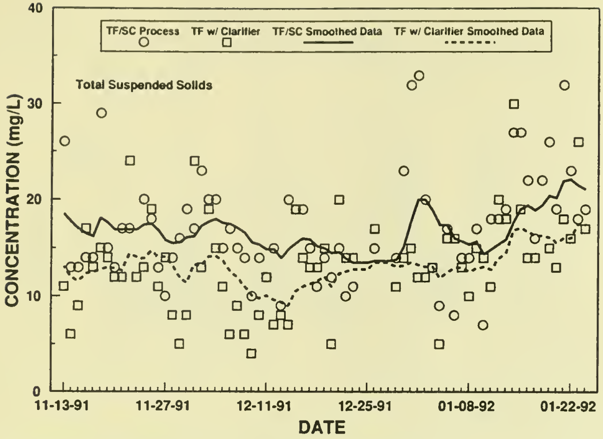


FIGURE 3.2 COMPARISON OF TF/SC AND TF WITH CLARIFIER PROCESS EFFLUENT CONCENTRATIONS

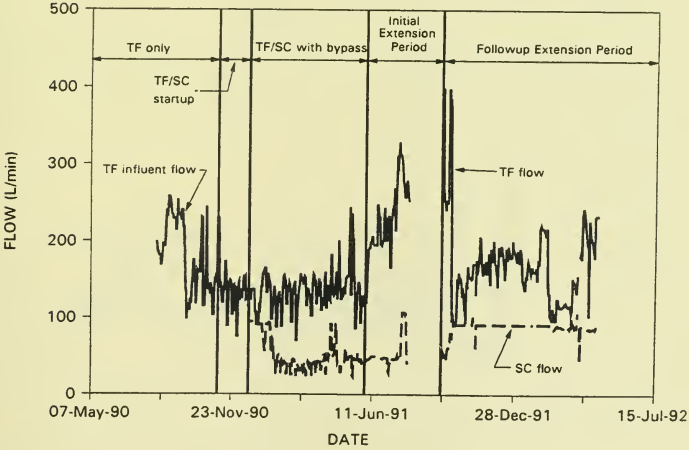


FIGURE 3.3 TF/SC AVERAGE DAILY FLOW

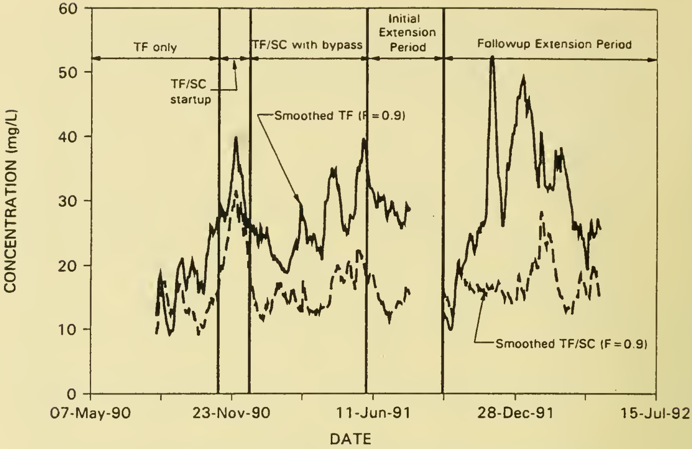


FIGURE 3.4 SMOOTHED TF EFFLUENT AND FINAL EFFLUENT SS CONCENTRATION

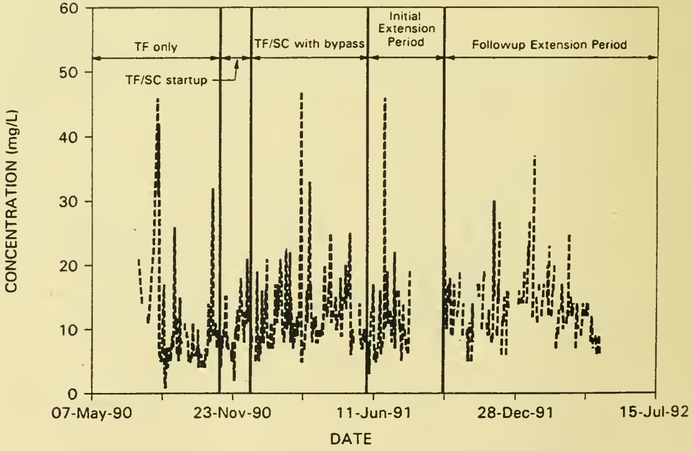


FIGURE 3.5 TF EFFLUENT SOLUBLE BOD_5 CONCENTRATION

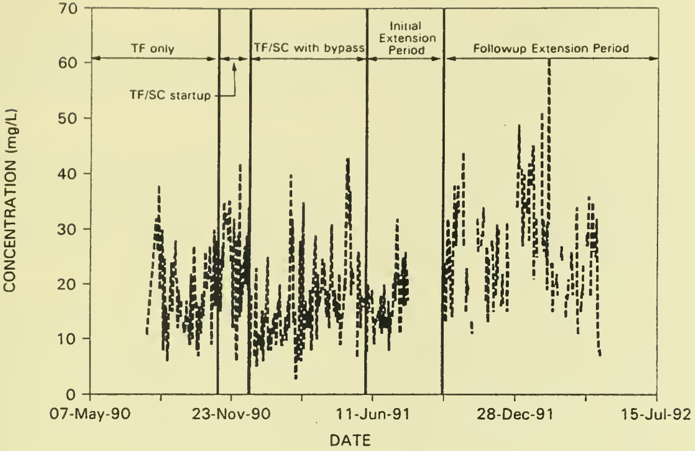


FIGURE 3.6 TF/SC FINAL EFFLUENT TOTAL BOD_5 CONCENTRATION

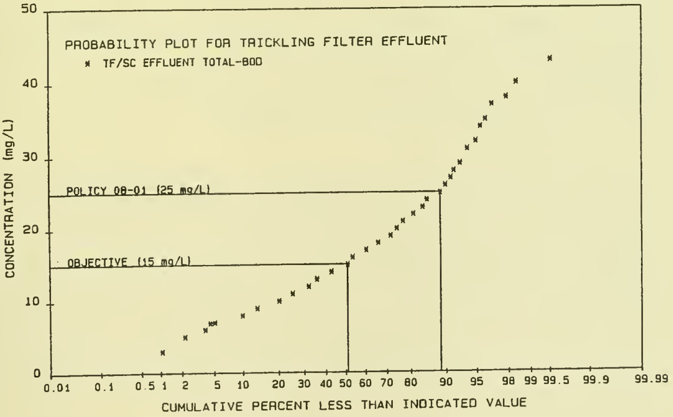


FIGURE 3.7 PROBABILITY PLOT OF TF/SC EFFLUENT TOTAL BOD_5 CONCENTRATION

concentration increased due to loading changes made to increase the stability of the SC unit. Figure 3-7 is a probability plot of the TF/SC effluent total BOD_5 concentration which shows that the concentration was less than 25 mg/L for 89 percent of the time during the first year of operation. The carbonaceous total BOD_5 concentration was on average about half that of the total BOD_5 concentration.

The smoothed trickling filter effluent and the final effluent total NH_3-N concentrations are shown in Figure 3-8. During the period when the TF was operated on its own, the trickling filter effluent and the final effluent ammonia concentrations were approximately the same. When the SC was started, the TF/SC or final effluent ammonia concentration was less than the TF effluent ammonia concentration. This indicates that nitrification was occurring in the SC tank during this winter period.

From December 18, 1990 to May 1, 1991, the TF/SC effluent concentrations were less than the effluent design objective concentrations on a daily basis for 55, 51, 64 and 100 percent of the time for SS, total BOD_5, TP, and un-ionized NH_3-N, respectively. The effluent NH_3-N concentration met the target of 2 mg/L for 82 percent of the time. It should be noted that total effluent NH_3-N was not an assigned objective for the plant, but was included as a quantitative measure of nitrification. These comparisons with the effluent design objective concentrations are for evaluation purposes only, since these assigned objectives refer not to single-sample requirements, but to monthly average design values. In terms of an upper limit value, the TF/SC effluent concentrations were compared with MOE Policy 08-01 limits for a secondary facility. These limits are 25, 25, and 1.0 mg/L for SS, total BOD_5 and TP. Compared to these limits, the effluent from the TF/SC was less than the limit for 92, 89, and 98 percent of the time on a daily basis.

During the follow-up extension period (September 17, 1991 to April 26, 1992) the effluent performance was not as good as the initial year of operation, because the focus during this period was to vary the operation of the system to try to establish a more effective SC.

TF/SC Loading Rates

Figure 3-9 is a plot of the TOL to the TF tower for the entire period. The loading varied from less than 0.1 to 1.3 kg $BOD_5/m^3 \cdot d$, averaging 0.42 kg $BOD_5/m^3 \cdot d$. The smoothed data show that the TOL varied throughout the period. Figure 3-10 is a trend plot of the TF wetting rate for the same period. Initially, a high wetting rate to the TF was used to assist in starting the unit. By late September 1990, the wetting rate was reduced to approximately 32 $m^3/m^2 \bullet d$ where it remained until the end of May 1991. The wetting rate was increased in June 1991 to evaluate operation of the TF during warmer temperatures and to assist in re-establishing the SC unit. The average wetting rate during this period was 55.5 $m^3/m^2 \bullet d$. During the follow-up extension period, the loading to the TF was generally increased to assist in maintaining a viable SC process. During this period the TOL and wetting rate averaged 0.47 kg $BOD_5/m^3 \bullet d$ and 41.5 $m^3/m^2 \bullet d$ respectively.

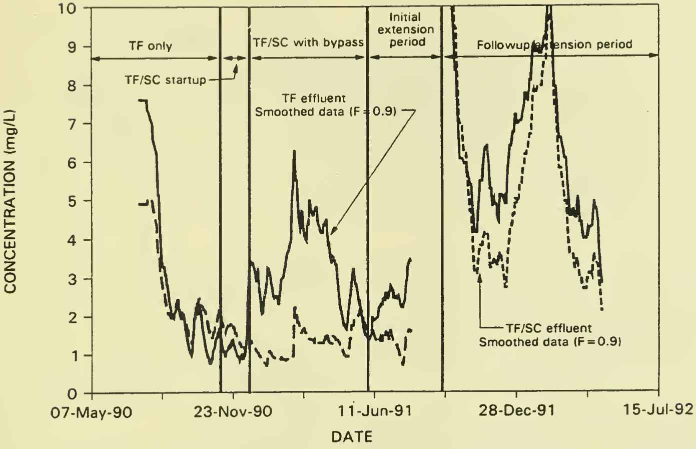


FIGURE 3.8 TF EFFLUENT AND FINAL EFFLUENT AMMONIA-N CONCENTRATION

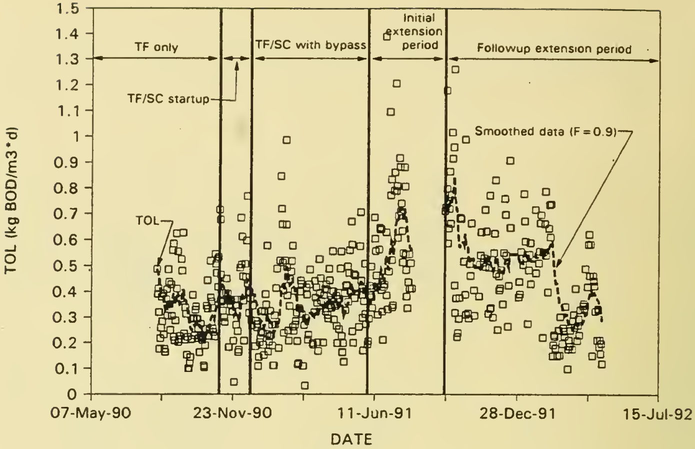


FIGURE 3.9 TF TOTAL ORGANIC LOADING (TOL) RATE

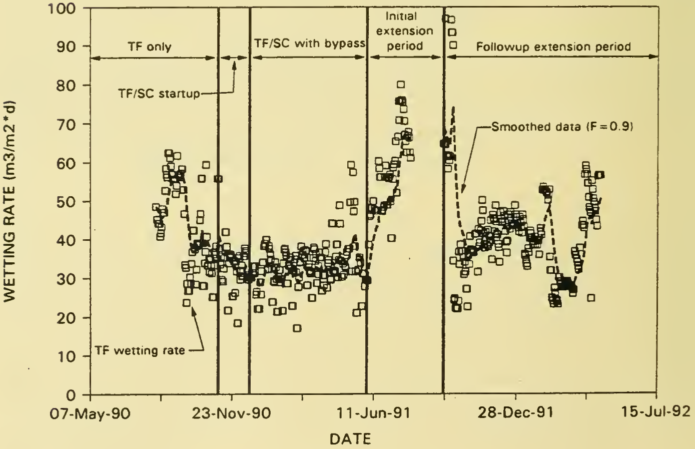


FIGURE 3.10 TF WETTING RATE

The SRT was calculated for the period when the SC was in operation. For this period, the SRT averaged 2.2 d and varied from less than 1 d to 7 d. The clarifier SOR trend is shown in Figure 3-11. Until December 18, 1990, all the flow treated by the TF was treated by the final clarifier. During this period, the SOR was greater than 30 $m^3/m^2 \cdot d$ and averaged 55 $m^3/m^2 \cdot d$. After December 18, the SOR was reduced to 30 $m^3/m^2 \cdot d$ and further reduced to 15 $m^3/m^2 \cdot d$ on January 11, 1991. During the follow-up extension period the SOR was maintained at a constant loading of about 28.0 $m^3/m^2 \cdot d$ to assist in reducing the HRT of the SC process.

Monitoring

The main monitoring parameters for the TF/SC are summarized in Table 3.9. The main TF/SC parameters were SVI, DO concentration, MLSS, and sludge blanket level. The SVI, for the period when the SC was operating, varied from 56 to 394 mL/g and averaged 151 mL/g. The wide range of SVIs was not expected and could not be associated with wasting practices, influent loading changes, or MLSS changes. The MLSS averaged 2973 mg/L after the SC unit was started in December 1990 and varied from 1000 to 5000 mg/L (Figure 3-12) once the SC was operating at near steady state conditions. Solids from the SC were lost late in April 1991. This loss is discussed further below.

The DO concentration varied from 2 to 11 mg/L; therefore, residual DO was available for biological activity at all times. The high DO concentrations were not associated with a shearing of the MLSS floc. Air flows used were required for adequate mixing of the suspended biomass. The sludge blanket level varied from 0.1 to 0.5 m and was the highest when the SC was in operation. When the SC was in operation, the sludge blanket averaged about 0.39 m.

Table 3.9
Summary of SC and Clarifier Monitoring Parameters

Period	SVI (mL/g)	DO Concentration (mg/L)	MLSS Concentration (mg/L)	Sludge Blanket Depth (m)
Overall Period When Operated* as TF/SC	151	5.6	2973	0.39
Trickling Filter Only (Aug. 30/90 to Nov. 4/90)	-	-	-	0.14
TF/SC Operation (Dec. 18/90 to May 1/91)	191	6.4	2810	0.37
Initial Extension Period - TF/SC (Jun. 1/91 to Aug. 2/91)	115	2.8	2640	0.41
Follow-up Extension Period - TF/SC (Sept. 17/91 to Apr. 26/92)	113	6.0	3288	0.40

Notes:
*Average of period when process operated as TF/SC process including:
Dec. 18/90 to May 1/91
Jun. 1/91 to Aug. 2/91
Sept., 17/91 to Apr. 26/92

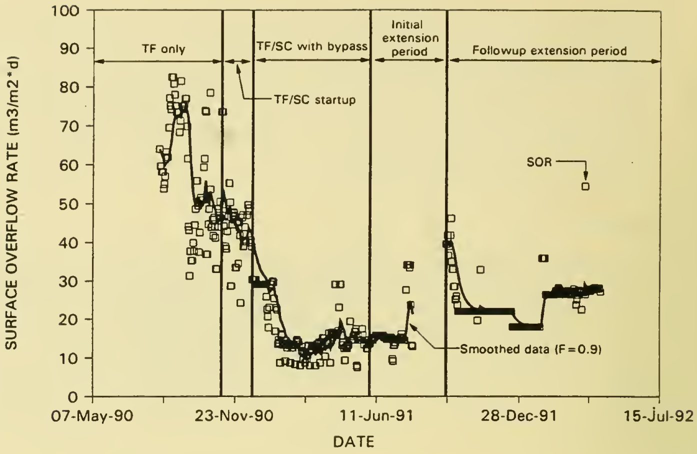


FIGURE 3.11 TF/SC FINAL CLARIFIER SURFACE OVERFLOW RATE

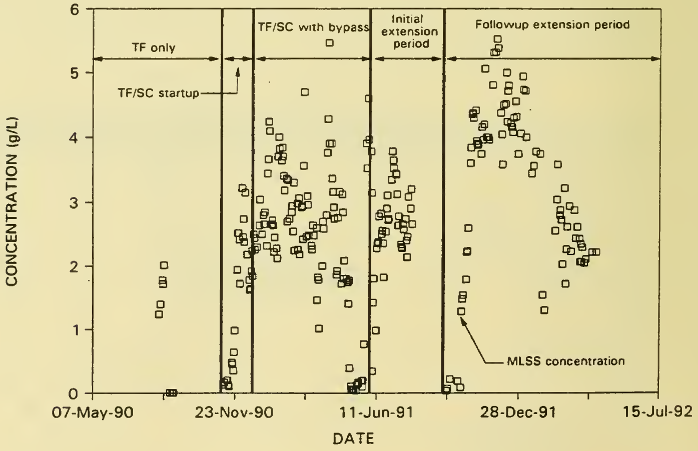


FIGURE 3.12 SOLIDS CONTACTOR MLSS CONCENTRATION

TF Nitrification

To investigate nitrogen removal through the TF tower, three sampling ports were installed along the height of the TF to sample for NH_3-N. The ports were located at one-half, one-third and three-quarters of the TF height. Figure 3-13 shows a number of profiles through the filter. The best-fit lines are based on an exponential equation and show some similarities between the profiles, especially the last three profiles that were taken. All the profiles show that the TF tower was removing ammonia throughout its depth.

Relationship Between Loadings and Response

The impact of the TOL on the effluent of the TF/SC was investigated. Figure 3-14 is a trend plot of the TOL and the effluent total BOD_5 concentration for the first year of operation. Figure 3-15 is a plot of the TOL versus the effluent total BOD_5 concentration for the same period. There is considerable scatter in the data and little evidence of a clear relationship. The relationship between solids contactor SVI and effluent SS concentrations was also investigated; however, there did not appear to be a correlation.

The TF nitrified throughout the entire testing period, including periods of colder wastewater temperatures. Figure 3-16 is a trend plot of the wastewater temperature and the TF effluent ammonia concentration for the first year of operation. The trend plot appears to show an increase in the TF effluent ammonia concentration around mid-December, which coincides with a period of reduced wastewater temperatures. There was also a decrease in the effluent ammonia concentration in mid-April as the wastewater temperature began to increase. Figure 3-17 shows a plot of the wastewater temperature versus the TF effluent ammonia concentration for the same period. The best-fit linear relationship ($r^2 = 0.2$) indicates that the TF effluent NH_3-N concentration cannot be related to the wastewater temperature.

SC Unit Solids Loss

Solids from the SC unit of the TF/SC process were lost at the beginning of May 1991. Immediately prior to the solids loss, the wastewater temperature had been increasing. Initially, it was felt that this wastewater temperature increase caused the TF to operate more efficiently, reducing the loading to the SC. Figure 3-18 shows the soluble F/M ratio for the SC unit from December 18, 1990 to April 26, 1992. The smoothed plot indicates that the F/M ratio did not change significantly during the test when start up periods are excluded. Since the soluble F/M ratio of the SC did not appear to change significantly at the end of April 1991, it is concluded that the SC was operating near its historical limit. The TF/SC was restarted in the fall of 1991 to investigate further the operation of the SC.

A solids balance was used to predict the solids inventory in the SC. The actual solids inventory accounts for the solids in the aeration basin and in the bottom of the second-

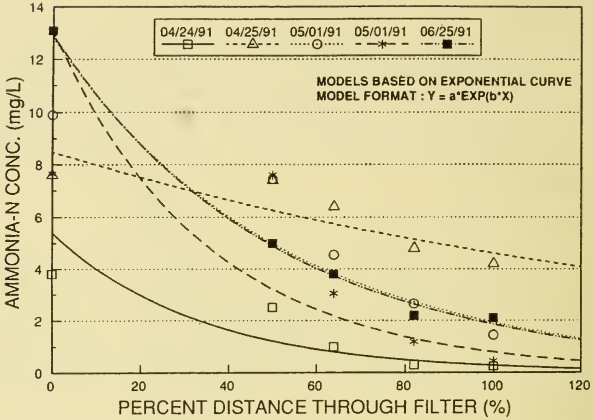


FIGURE 3.13- AMMONIA-N CONCENTRATION PROFILES THROUGH TRICKLING FILTER TOWER

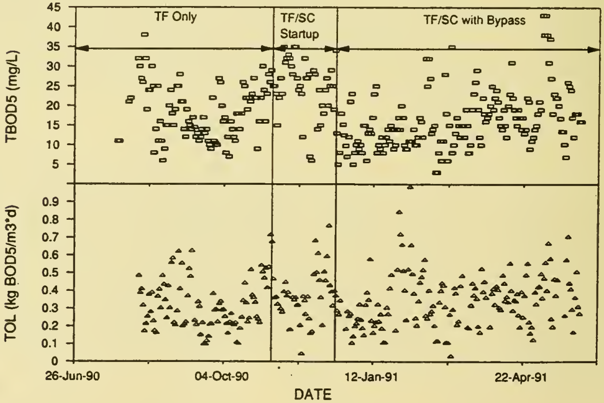


FIGURE 3.14 TREND PLOT OF TOL AND FINAL EFFLUENT TOTAL BOD_5 CONCENTRATION

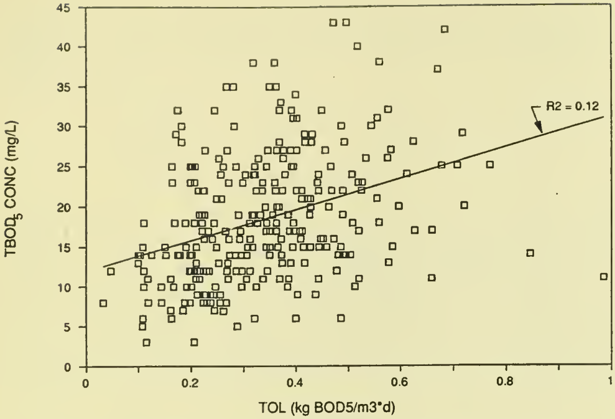


FIGURE 3.15 TOL VERSUS FINAL EFFLUENT TOTAL BOD_5 CONCENTRATION

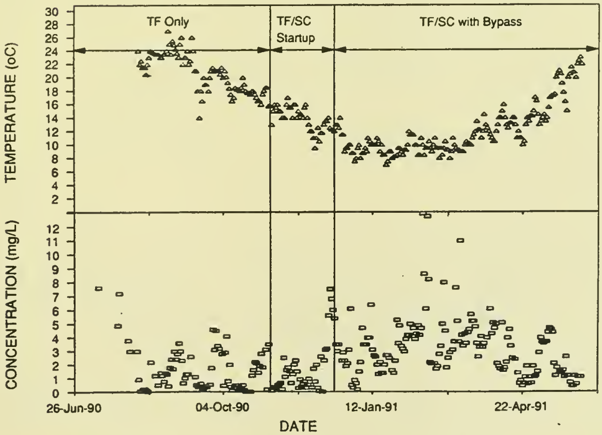


FIGURE 3.16 TREND PLOT OF WASTEWATER TEMPERATURE AND TF EFFLUENT AMMONIA-N CONCENTRATION

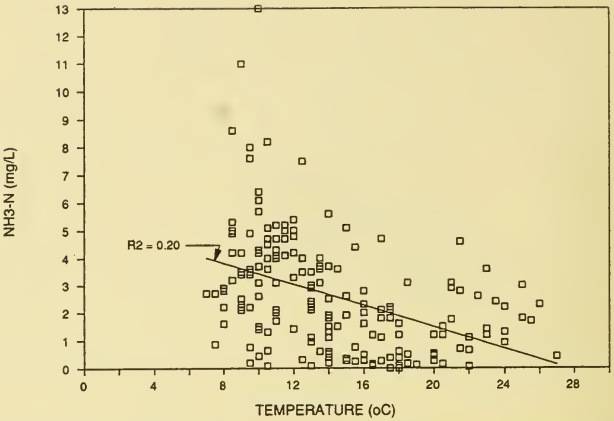


FIGURE 3.17 WASTEWATER TEMPERATURE VERSUS TF EFFLUENT AMMONIA-N CONCENTRATION

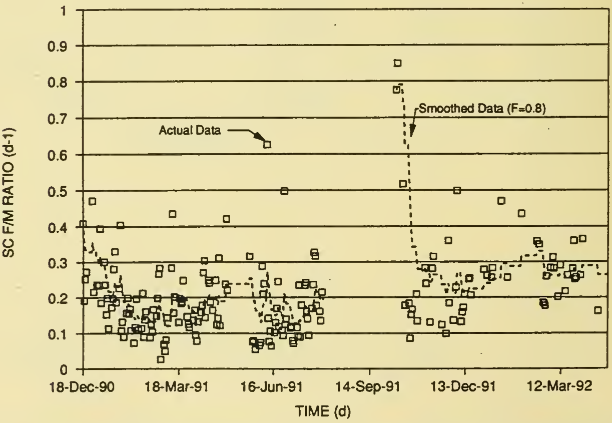


FIGURE 3.18 SOLUBLE BOD_5 F/M RATIO OF SOLIDS CONTACTOR

ary clarifier. Several assumptions were necessary to calculate the balance. The main assumption was that the solids in the final clarifier are the average of the MLSS and RAS concentrations. Figure 3-19 shows the actual and predicted solids inventory. The predicted solids inventory is based on:

- the difference in the solids entering the SC and the solids in the final effluent
- the biomass yield from the soluble BOD_5 utilized
- the solids wasted from the system.

The predicted solids inventory tracks the actual solids inventory until the solids loss in May 1991. The trend shows a gradual reduction in the solids inventory from December 1990 to May 1991. The solids inventory was approximately 20 kg in December 1990 and decreased to 10 kg in March 1991. To start the SC unit, the tank was seeded on two occasions during the first week in November with approximately 1 m^3 of RAS from the Little River PCP. Both solids inventories indicate a decrease in solids at the end of April 1991. After the solids were lost from the SC at the beginning of May, the predicted solids inventory was reset to zero. After the unit was restarted, the predicted inventory was significantly greater than the actual inventory. During this period, there were some problems with the return pump and these problems may have caused a solids loss which is not evident in the monitoring data. Further, the solids balance is based strictly on a steady state balance. It should be noted that throughout the period, the gradual solids loss was not reflected in decreasing MLSS concentration; therefore, it is believed that the clarifier sludge blanket was critical to process operation.

During the follow-up extension period a number of operational changes were attempted to improve the operation of the SC process, including:

- increased hydraulic and organic loading of TF
- reduction of SC hydraulic retention time
- operation of the TF/SC in a mode III configuration with a solids reaeration tank

None of these actions appeared to improve the viability of the SC or improve effluent quality.

Solids Yield

Wasting from the TF/SC was accomplished either by wasting from the SC or from the RAS line. Figure 3-20 shows the cumulative intentional and unintentional solids lost and total BOD_5 removed from December 18, 1990 to May 1, 1991. The solids lost from the system are the solids intentionally wasted and the solids unintentionally lost in the effluent. The BOD_5 removed is the difference between the primary influent and final effluent total BOD_5 loading. The figure shows that the amount of solids that is wasted intentionally from the system and the amount of solids that is unintentionally lost from the system in the effluent is approximately the same. Over this period, the overall solids yield was 1.10 kg TSS/kg BOD_5 removed or 0.79 kg VSS/kg BOD_5 removed.

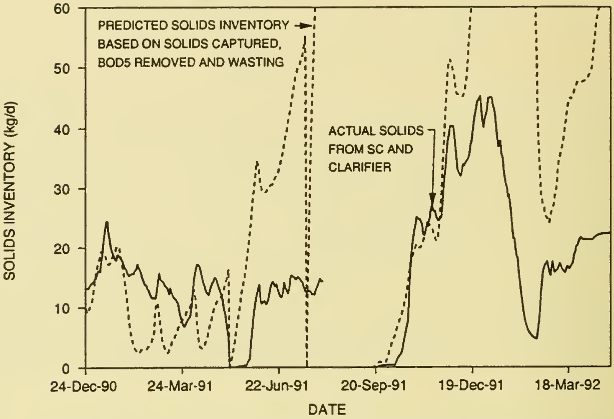


FIGURE 3.19 ACTUAL AND PREDICTED SOLIDS INVENTORY IN TF/SC SUSPENDED GROWTH SYSTEM

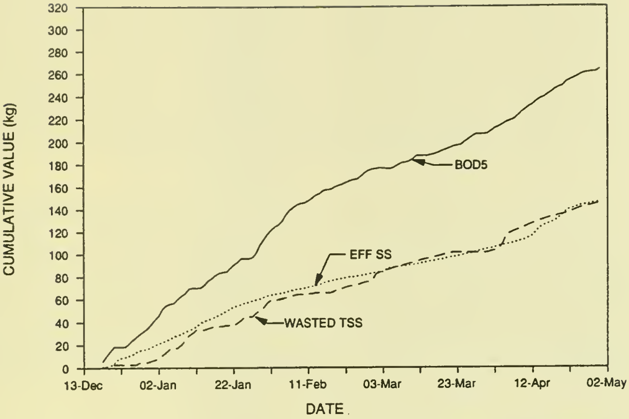


FIGURE 3.20 CUMULATIVE SOLIDS WASTED AND TOTAL BOD_5 REMOVED FOR TF/SC PROCESS

RBC PROCESS

General

The RBC pilot plant began operation on June 26, 1990 and the unit was operated continuously until the end of May 1991. Table 3.10 summarizes the influent parameters (wastewater flow, pH, and temperature) and the effluent concentrations for four periods. The four periods correspond to the overall period and the three different periods of varying influent concentrations due to the operation of the two clarifiers feeding the pilot plants. During the first period, both clarifiers operated with alum and polymer. One clarifier operated only with alum during the second period. During the third period, both clarifiers operated with ferric chloride and polymer. The overall period is based on more than eleven months of continuous operation.

The results indicate that the RBC pilot plant was not significantly affected by changes in the influent characteristics. There was an increase in the SS, total BOD_5 and TP during the period when polymer was shut off from the clarifiers feeding the pilot plants; however, the increase is less than 20 percent for the overall period. On average, the RBC met the effluent objectives for SS, total BOD_5, TP, and un-ionized NH_3-N for every period. The process target total NH_3-N concentration of 2.0 mg/L was exceeded during the winter period (period 3) due to intentionally sustained high hydraulic and nitrogenous loadings. The RBC did meet the reassessed total NH_3-N concentration of 3 mg/L, on an average period basis.

Hydraulic and Process Loadings

The operating parameters for the RBC for the same periods discussed above are shown in Table 3.11. The RBC treated within 95 percent of its initial target hydraulic loading of 88 $m^3/m^2 \cdot d$. On a peak day basis, the RBC treated a maximum daily flow of 205 $m^3/m^2 \cdot d$. This flow relates to a 2.4 peaking factor above the average flow treated by the unit or 2.3 times the initial target average loading of 88 $m^3/m^2 \cdot d$. The unit was operated continuously five days a week using a simplified diurnal flow pattern modelled after the diurnal flow entering the full-scale plant.

The RBC operated close to its initial target loading rates for organic, ammonia, and hydraulic loadings. The organic loading slightly exceeded the target of 3.2 g $BOD_5/m^2 \cdot d$, and was on average 3.5 g $BOD_5/m^2 \cdot d$. Both the ammonia and the RBC hydraulic loadings were slightly less than the initial target values of 1.3 g $TKN/m^2 \cdot d$ and 88 $m^3/m^2 \cdot d$. The final clarifier SOR was maintained at an average of 31 $m^3/m^2 \cdot d$, with a peak daily loading of 75 $m^3/m^2 \cdot d$. These values are twice the expected hydraulic loading rate based on the review of the process and the design of conventional secondary clarifiers for suspended growth systems. However, at this SOR, the effluent SS concentration was on average less than 10 mg/L and 15 mg/L for 85 and 95 percent of the time, respectively.

Table 3.10
Summary of Performance of the RBC Pilot Plant

Period	Influent Parameters			Effluent Concentration (mg/L)					
	Flow (L/min)	pH	Temp (°C)	SS	Total BOD_5	Carbonaceous BOD_5	TP	Un-ionized NH_3-N	Total NH_3-N
Overall Period (Jul. 5/90 to May 31/91)	43	7.1	16	8	11	7	0.3	0.014	2.1
Alum + Polymer Addition (Jul. 5/90 to Oct. 29/90)	50	7.2	22	7	12	7	0.3	0.018	1.6
Alum Only (Oct. 30/90 to Jan. 10/91)	41	7.1	15	9	14	8	0.4	0.013	2.0
Ferric and Polymer Addition (Jan. 11/91 to May 31/91)	39	7.1	13	7	10	6	0.2	0.013	2.6

Table 3.11
Summary of RBC Operating and Monitoring Parameters

Period	Operating Parameters					Monitoring
	Organic Loading (g $BOD_5/m^2 \cdot d$)	Nitrogen Loading (g $TKN/m^2 \cdot d$)	Hydraulic Loading ($m^3/m^2 \cdot d$)	HRT (h)	Clarifier SOR ($m^3/m^2 \cdot d$)	Average DO* Concentration (mg/L)
Overall Period (July 5/90 to May 31/91)	3.5	1.2	84.1	1.5	30.7	3.5
Polymer Addition (Jul. 5/90 to Oct. 29/90)	3.0	1.2	99.1	1.3	36.2	2.8
Without Polymer Addition (Oct. 30/90 to Jan. 10/91)	3.8	1.2	81.4	1.5	29.7	4.0
Ferric and Polymer Addition (Jan. 11/91 to May 31/91)	3.7	1.2	76.7	1.7	28.0	3.7

Notes:
*Average DO concentration is average of DO measurements taken in first and last stages.

Trends

Figure 3-21 is a trend plot of the RBC effluent showing the effluent SS concentration. The plot shows that the spread of effluent SS concentrations is generally between 4 and 12 mg/L. Figure 3-22 is a probability graph of the same period, which shows that the effluent SS concentration was less than 15 mg/L for 95 percent of the time on a daily basis.

Figure 3-23 is a trend plot of the effluent TP concentrations that shows that the effluent concentration generally ranged from 0.2 to 0.4 mg/L. Figure 3-24 is a trend plot of the effluent total NH_3-N concentration, showing that it was above 2 mg/L for a significant period of the time. The higher total NH_3-N concentrations occurred on a number of occasions and did not appear to be linked only to low wastewater temperatures; these concentrations are explored in further detail below. Despite higher effluent total NH_3-N concentrations, the un-ionized ammonia concentration was less than 0.1 mg/L for the entire period.

From July 5, 1990 to May 31, 1991 the effluent concentrations from the RBC were less than the objective concentrations on a daily basis for 96, 81, 92, and 100 percent of the time for SS, total BOD_5, TP and un-ionized NH_3-N. The effluent total NH_3-N concentration met the target of 2 mg/L for 55 percent of the time. In terms of an effluent ammonia-N concentration of 3 mg/L, used during the follow-up extension, the RBC effluent met this limit 74 percent of the time. These comparisons with the effluent objective concentrations are for evaluation purposes only, since these objectives did not refer to single-sample requirements, but to monthly average values. In terms of an upper limit the effluent parameters from the RBC were compared with Policy 08-01 limits for a secondary facility. These limits are 25, 25, and 1.0 mg/L for SS, total BOD_5 and TP. The RBC effluent was less than these limits for 99, 99, and 98 percent of the time.

The process loadings through the RBC did not vary significantly over the testing period. The organic loading rate was well distributed around the target loading of 3.2 g $BOD_5/m^2{\bullet}d$ and averaged 3.5 g $BOD_5/m^2{\bullet}d$. The loading was less than the objective during the first part of the testing when the clarifiers feeding the pilot plants were using alum and polymer for particulate removal. When the polymer was shut off or the plant was using ferric, the organic loading increased to between 3.7 and 3.8 g $BOD_5/m^2{\bullet}d$. Loading increased during the period (Jan 11 to May 31, 1991) when the plant was using ferric and polymer because of an increased soluble BOD_5 concentration and a reduction in the polymer dosage to clarifiers No. 5 and 6. Figure 3-25 shows an actual and smoothed trend plot of the organic loading during the testing. Smooth data is based on a 7-d effective value. The ammonia loading varied from 0.6 to 2.2 g $TKN/m^2{\bullet}d$ and averaged 1.2 g $TKN/m^2{\bullet}d$. The hydraulic loading rate varied from 30 to 210 $m^3/m^2{\bullet}d$ and averaged 84 $m^3/m^2{\bullet}d$. A number of higher hydraulic loading rates were evident in the fall and spring and are associated with peak flow events. Figure 3-26 shows the actual and smoothed trend plot of the hydraulic loading rate for the testing period. The SOR of the final clarifier averaged 31 $m^3/m^2{\bullet}d$, which is considered near the peak

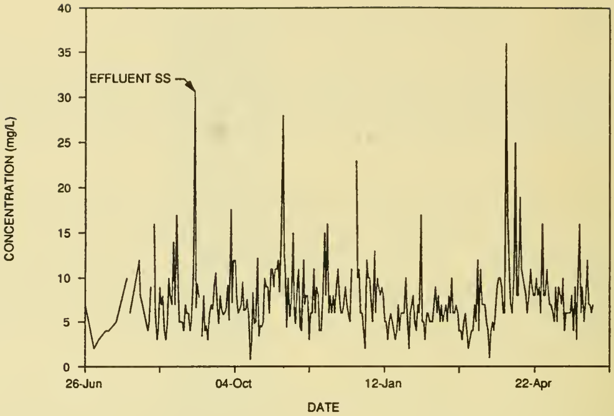


FIGURE 3.21 RBC EFFLUENT SS CONCENTRATION

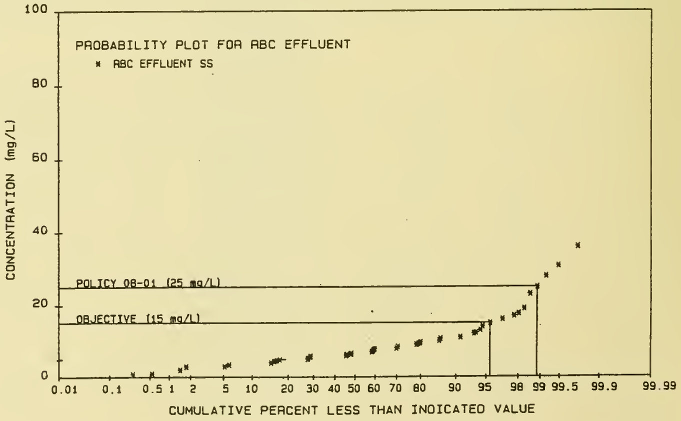


FIGURE 3.22 PROBABILITY PLOT OF RBC EFFLUENT SS CONCENTRATION

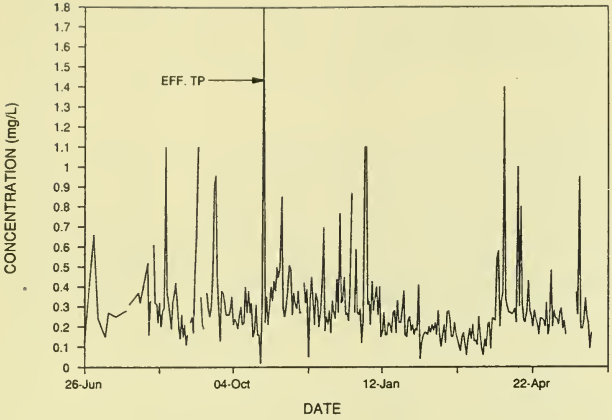


FIGURE 3.23 RBC EFFLUENT TP CONCENTRATION

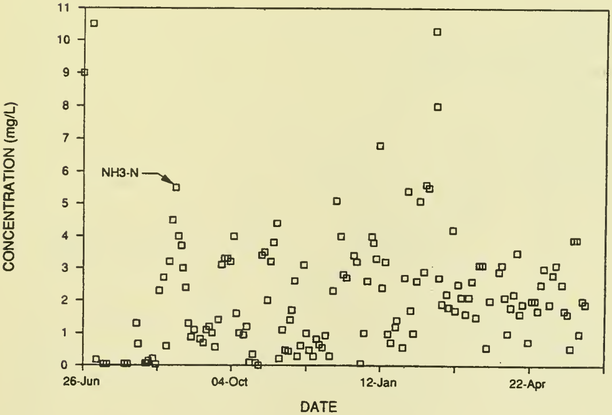


FIGURE 3.24 RBC EFFLUENT AMMONIA-N CONCENTRATION

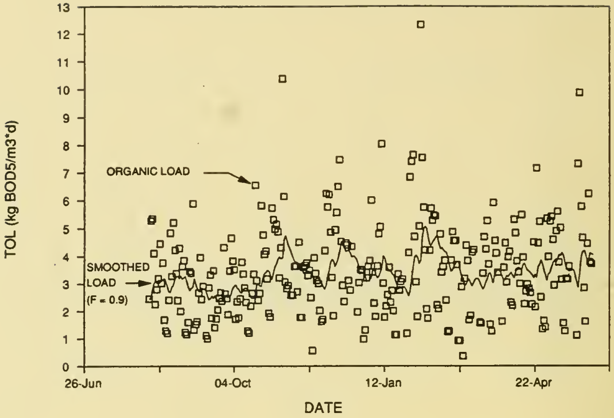


FIGURE 3.25 TREND PLOT OF RBC TOTAL ORGANIC LOADING (TOL)

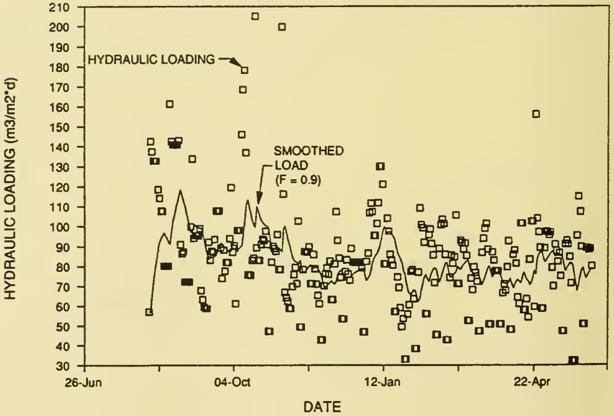


FIGURE 3.26 TREND PLOT OF RBC HYDRAULIC LOADING

day design SOR for final clarifiers. The RBC final clarifier handled a SOR of up to 75 $m^3/m^2 \cdot d$, with no apparent impact on the pilot plant's effluent quality.

Relationship Between Loadings and Response

It was expected that given the increases in the effluent NH_3-N concentration from the RBC during the winter, a relationship between the wastewater temperature and the effluent total NH_3-N concentration would exist. Figure 3-27 is a trend plot of the wastewater temperature and the effluent total NH_3-N concentration. There appears to be an increase in the effluent NH_3-N concentration beginning mid-December and proceeding until early March; however, a linear relationship was not significant. The impact of increased organic loading on the RBC effluent quality was also investigated. Figure 3-28 is a plot of the organic loading versus the effluent total BOD_5 concentration. The trend ($r^2 = 0.2$) indicates an apparent increase in the effluent total BOD_5 concentrations and increasing organic loading to the RBC; however, the trend cannot confirm this observation.

NH_3-N removal through the RBC was determined by taking discrete samples at the end of each stage. Figure 3-29 shows the total NH_3-N concentration profile through the RBC for a number of days; each profile was fitted with an exponential equation. The retention time is based on the total liquid volume divided by the average daily flow rate, and this time is divided equally between each sampling point. The plots show that NH_3-N removal occurs throughout the RBC unit.

Solids Yield

Wasting from the RBC was accomplished by timed operation of a pump wasting from the final clarifier. Figure 3-30 is the cumulative solids generated, solids lost, and total BOD_5 removed for the period beginning June 26, 1990. The solids generated are the solids settled in the clarifier and wasted. The solids lost from the system are the solids unintentionally lost in the effluent. The total BOD_5 removed is the difference between the influent and effluent total BOD_5 loading. The figure shows that about twice the amount of solids are wasted intentionally from the system than the amount that is lost from the system in the effluent. Over this period, the overall solids yield was 0.99 kg TSS/kg BOD_5 removed, or 0.78 kg VSS/kg BOD_5 removed.

BAF PROCESS

General

The BAF pilot plant began continuous operation on August 20, 1990 and was operated until August 14, 1991. The BAF pilot plant was restarted on September 17, 1991 and operated until April 26, 1992 to resolve outstanding issues with respect to its operation. Table 3.12 summarizes the influent parameters (flow, pH, and temperature) and effluent concentrations for six periods. The six periods correspond to the overall period, the three periods of full-scale plant operation, the initial extension period and

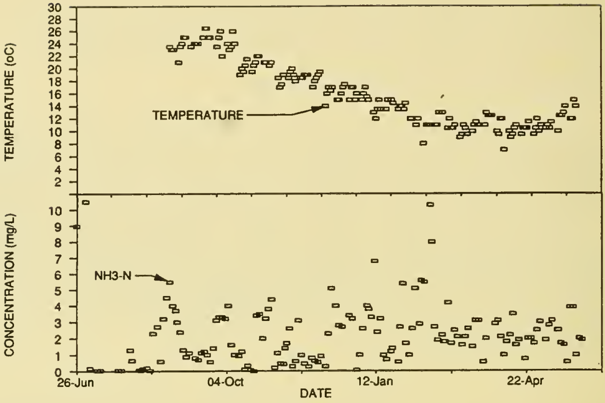


FIGURE 3.27 TREND PLOT OF WASTEWATER TEMPERATURE AND RBC EFFLUENT AMMONIA-N CONCENTRATION

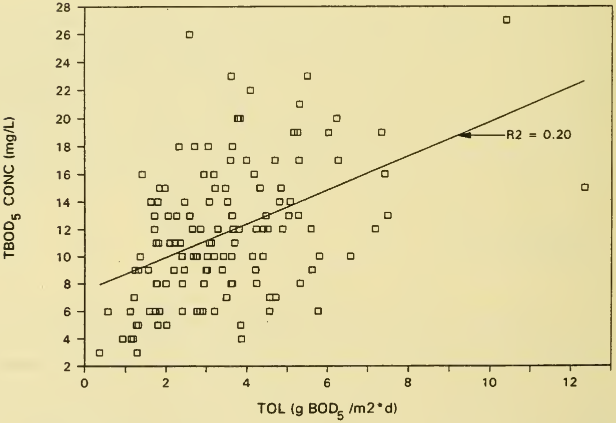


FIGURE 3.28 RBC TOTAL ORGANIC LOADING (TOL) VERSUS EFFLUENT TOTAL BOD_5 CONCENTRATION

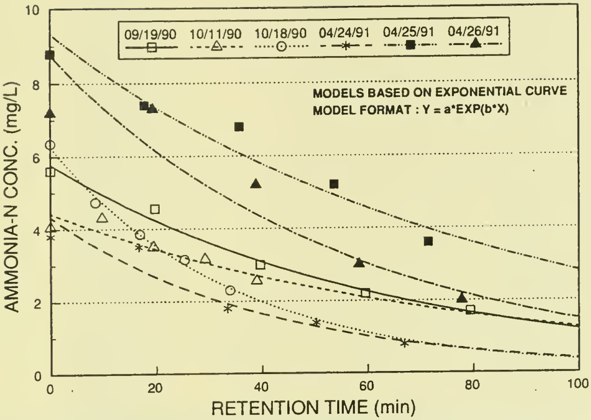


FIGURE 3.29 - AMMONIA-N CONCENTRATION PROFILES THROUGH RBC TANK

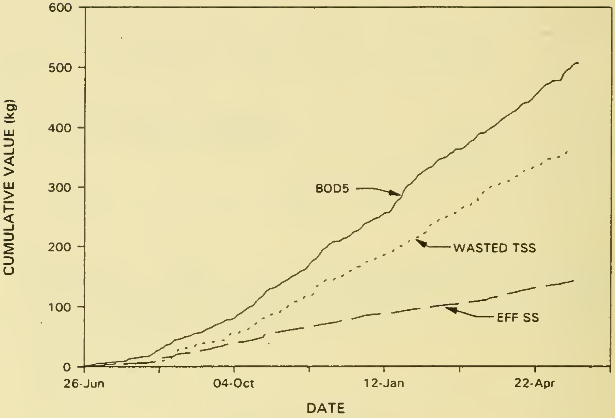


FIGURE 3.30 CUMULATIVE SOLIDS WASTED AND TOTAL BOD_5 REMOVED FOR RBC PROCESS

Table 3.12
Summary of Performance of the BAF Pilot Plant

Period	Influent Parameters			Effluent Concentrations (mg/L)					
	Flow (L/min)	pH	Temp (°C)	SS	Total BOD_5	Carb. BOD_5	TP	NH_3-N	Un-ionized NH_3-N
Overall Period (Aug. 20/90 to Apr. 26/92, excluding Jan. 11/91 to Mar. 11/91)	19	7.2	16	5	10	7	0.3	5.0	0.032
Alum plus Polymer Addition (Aug. 20/90 to Oct. 29/90)**	23	7.1	21	4	8	5	0.1	1.9	0.016
Alum Without Polymer Addition (Oct. 30./90 to Jan. 10/91)**	23	7.7	14	7	13	9	0.3	6.2	0.037
Ferric and Polymer Addition (Jan. 11/91 to May 31/91)**	15	7.5	13	4	10	7	0.1	8.1	0.052
Initial Extension Period (Jun. 1/91 to Aug. 2/91)	17	7.1	21	4	10	6	0.4	3.1	0.027
Follow Up Extension Period (Sept. 17/91 to Apr. 26/92)	18	7.1	13	4	9	5	0.3	5.1	0.028
Increased Media Depth (2.5 m) (Jan. 15/92 to Apr. 26/92)*	20	7.1	11	4	8	4	0.4	3.0	0.011

Notes:
*Elevated TP during extension period due to high influent concentration
**Polymer not expected to be a factor, nitrification affected by operation of pilot plant; i.e. vigorous backwashes

the follow-up extension period. The three periods of full-scale plant operation refer to the use of alum and polymer, alum only, and ferric and polymer during the pilot testing period. The initial extension was conducted from June 1 to August 2, 1991 to show that nitrification could be re-established and to simulate winter operations; however, attempts were not successful partly due to equipment limitations. The follow-up extension was conducted from September 17, 1991 to April 26, 1992 and was initiated to resolve outstanding issues with respect to nitrification and hydraulics.

During the follow-up extension period, the media depth was increased from 2.0 to 2.5 m starting on January 15, 1992. The average performance for this period has also been shown in Table 3.12.

Nitrification was lost from the BAF in mid-December 1990, and attempts to re-establish nitrification from January 11 to March 11, 1991 were unsuccessful. Nitrification was lost due to a number of vigorous backwashes that were intentionally conducted in November and December 1990. These backwashes were conducted to clean the media bed, since the unit was experiencing apparent hydraulic bottlenecks. The initial extension period was conducted to show that the nitrifiers would be re-established in the unit as the wastewater temperatures increased late in the spring. The extension also attempted to simulate winter wastewater temperatures by chilling the influent wastewater. Unfortunately, with raw wastewater temperatures of 25°C, attempts to bring the BAF temperature to 10°C were unsuccessful. The unit was restarted on September 17, 1991 and operated until April 26, 1992 to ensure the unit could nitrify at cold wastewater temperatures and could handle peak-day hydraulic events.

Effluent Quality

Table 3.12 indicates that the pilot plant treated an average flow rate of 19 L/min or 2.3 $m^3/m^2 \cdot h$, which is close to the initial target flow of 21 L/min or 2.5 $m^3/m^2 \cdot h$. The average effluent quality was excellent for SS and TP averaging 5 and 0.3 mg/L for the overall period. The effluent TP concentration increased near the end of June 1991, due to an increase in the influent TP concentration of 15 to 30 mg/L over a two-day period. The total and carbonaceous BOD_5 concentrations averaged 10 and 7 mg/L respectively.

During the first period (August 20 to October 19, 1991), the ammonia concentration averaged 1.9 mg/L, and both the total and carbonaceous BOD_5 concentrations were less than 8 mg/L. For the overall period, the average effluent total NH_3-N concentration averaged 5.0 mg/L because the unit lost nitrification during the winter of 1991 for the reasons discussed above. The total BOD_5 concentration was greater than the carbonaceous concentration, since the unit was not fully nitrifying for most of the period. The effluent un-ionized NH_3-N averaged 0.03 mg/L which met the design objective of 0.1 mg/L.

During the follow-up extension period (September 17, 1991 to April 26, 1992) the BAFs nitrification performance continued to be marginal, until the media depth was increased from 2.0 to 2.5 m on January 15, 1992. Initially, the influent phosphorus was augmented, since it was thought that the biological process was nutrient limited. Table 3.13 summarizes the period with and without phosphorus augmenting. Indications are that the consistent nitrification was due to the increased bed depth and not available phosphorus. However, more ortho-phosphorus analysis would be required to confirm this observation.

Table 3.13
Comparison of BAF Pilot Plant Performance With and Without Phosphorus Augmenting

Period	Effluent Concentration (mg/L)				
	TSS	$TBOD_5$	$CBOD_5$	TP	NH_3-N
With Phosphorus Augmenting (Jan. 15/92 to Feb. 28/92)	5	8	4	0.8*	3.2
Without Phosphorus Augmenting (Mar. 1/92 to Mar. 22/92)	4	9	6	0.16	2.6

Notes: * Elevated TP concentration due to augmentation

However, phosphorus for nutrient removal may be limiting due to the operation of the full-scale plant. The phosphorus concentration will have to be closely monitored for any of the secondary processes if the City intends to maintain current full-scale plant operations.

It is difficult to determine the impact of the period when polymer was removed from the full-scale plant, since the unit lost nitrification during this same period due to the operational activity described earlier. However, the SS, carbonaceous BOD_5, and TP concentrations did increase during this period due to increased organic and particulate loadings (Table 3.12). Organic loading increased by 50 percent during the period without polymer. The effluent SS averaged 4 mg/L when alum and polymer were used in the full-scale plant (August 20 to October 29, 1990), and increased to 7 mg/L when the polymer was shut off (October 30, 1990 to January 10, 1992). Similarly, the carbonaceous BOD_5 concentration increased from 5 to 9 mg/L over the same two periods. The TP also increased from 0.1 to 0.3 mg/L. Despite the increases in the effluent concentrations with or without polymer used in the full-scale plant, the BAF obtained excellent removal of particulates and associated contaminants.

The key operating and monitoring parameters for the BAF are shown in Table 3.14. Overall, the BAF was loaded at the initial target values for organic and hydraulic loading. The ammonia loading, set at a maximum of 0.7 g $TKN/m^3 \cdot d$, was based on a wastewater temperature of 15°C. On average, the total NH_3-N loading was about half of the TKN design loading and nitrification was evident until it was disrupted by the pilot plant operations discussed above.

Table 3.14
Summary of BAF Operating and Monitoring Parameters

Period	Operating Parameters				Monitoring Parameters	
	Organic Loading (kg $BOD_5/m^3 \cdot d$)	Nitrogen Loading (kg $TKN/m^3 \cdot d$)	Hydraulic Loading ($m^3/m^2 \cdot h$)	HRT (min)	DO (mg/L)	Air Flow (cfm)
Overall Period (Aug. 20/90 to Apr. 26/92, excluding Jan. 11 to Mar. 11/91)	1.19	0.39	2.30	62.2	3.6	6.9
Alum plus Polymer Addition (Aug. 20/90 - Oct. 29/90)	1.04	0.41	2.82	47.8	3.2	6.8
Alum Without Polymer Addition (Oct. 30/90 to Jan. 10/91)	1.50	0.49	2.81	47.7	4.7	7.6
Ferric and Polymer Addition (Jan. 11/91 to May 31/91)	0.99	0.32	1.79	87.6	3.8	4.9
Initial Extension Period (June 1/91 to Aug. 2/91)	1.10	0.35	2.05	75.4	1.9	7.1
Follow Up Extension Period (Sept. 17/91 to Apr. 26/92)	1.10	0.35	2.13	68.7	2.8	7.0
Increased Media Depth (2.5 m) (Jan. 15/92 to Apr. 26/92)	0.95	0.33	2.41	69.8	5.1	7.7

The BAF effluent consistently met the objectives for all parameters except for the effluent total NH_3-N concentration. From August 20, 1990 to May 31, 1991, excluding the period of low flows when attempts were made to re-establish nitrification (January 11 to March 11, 1991), the effluent concentrations were less than the effluent design objective concentrations on a daily basis for 98, 82, 96, and 100 percent of time for SS, total BOD_5, TP and un-ionized NH_3-N. For this period, the effluent total NH_3-N concentration was less than the objective of 2 mg/L, for 22 percent of the time. If the initial extension period from June to July 1991 is used, the BAF effluent met the assigned total NH_3-N target for 35 percent of the time. During this time the unit may not have reached a steady nitrification rate. During the follow-up extension period after the media depth had been increased to 2.5 m, the BAF effluent met the assigned NH_3-N target of 3 mg/L for 40 percent of the time.

If the period prior to disruption of nitrification (September 7 to October 31, 1990) and the period of increased media depth (January 28, to April 26, 1992) are evaluated, the effluent NH_3-N concentration was less than 2 mg/L for 53 percent of the time. If an effluent NH_3-N target of 3 mg/L is used, the BAF process achieved less than the target for 71 percent of the time for the same period.

These comparisons with the objective concentrations are for comparative purposes only, since the effluent design objective concentrations do not refer to single-sample values but to monthly average values.

In terms of an upper limit value, the effluent concentrations from the BAF were compared with the MOE Policy 08-01 limits for SS, total BOD_5, and TP, which are 25, 25, and 1.0 mg/L, respectively. Based on these upper limits, the BAF effluent was less than these limits for 99, 98, and 99.5 percent of the time.

Hydraulic Loading

In terms of the capability of the BAF to handle peak day flows, there were initial indications that the unit had physical limitations treating the 2.2 peak factor. Overall, the unit did treat a 2.0 peaking factor, and was able to handle in excess of a 2.2 peaking factor for a day-long period using tests with clean water on May 23, 1991. A clean water hydraulic test was conducted because occasionally the filter appeared to clog during the high flow periods, which caused it to syphon. A full-scale system would have an automatic backwashing provision to remedy this problem. The pilot plant also had difficulty handling the diurnal flow pattern during the study period, which resulted because the unit syphoned after the pilot plant "air bridged or air locked". The BAF pilot plant appeared to "air lock" due to partial clogging and due to the small surface area of the pilot plant. The supplier indicated that in a full-scale plant, partial clogging can occur which causes poor air distribution. This clogging was corrected by conducting an air scouring once a day to loosen the filter. The pilot scale unit had a surface area of only 0.5 m^2, whereas full-scale units can have surface areas of 84 m^2. Therefore, the wall effects in the small pilot unit may have been a problem. Air scouring was used in July 1991 and the "air locking" problem appeared to be corrected.

Remaining hydraulic concerns were addressed during the follow-up extension period when two hydraulic tests were conducted for a 3- to 4-d period. The second hydraulic test, shown in Figure 3-31, was conducted at a hydraulic loading of 5.5 m^3/m^2•h for a 3-d period. The trend plot shows the unitized flow rate and backwash events. The design average hydraulic loading was 2.5 m^3/m^2•h; therefore, a 2.2 peaking factor is 5.5 m^3/m^2•h. The backwash events are displayed as the number of cycles per event. An air scouring event is referenced as a negative backwash event. Although the number of backwash events increased, the percentage of backwash volume to treated flow was similar to normal operations averaging about 13 percent.

Trends

Figure 3-32 is a trend plot of the treated flow rate. The flow rate varied considerably for the first three months and averaged 23 L/min. The period from January 11 to March 11, 1991, when attempts were made to re-establish nitrification in the BAF, shows as a period of lower flow on the trend plot. After March 11, 1991, the flow was increased to ascertain performance regardless of the degree of nitrification and averaged around 21 L/min or 2.5 m^3/m^2•h. During the follow-up extension period, the hydraulic loading averaged 18 L/min (2.2 m^3/m^2•h) although once nitrification was established with a greater media depth, the flow rate through the unit averaged 20 L/min or 2.4 m^3/m^2•h.

Plots of the effluent SS and TP show the excellent effluent quality obtained from this unit, especially in terms of particulate and associated contaminant removal. Figure 3-33 shows the BAF effluent SS concentration, which was less than 5 mg/L for the majority of the time. The effluent SS concentration from the BAF was less than 5 mg/L, for 60 percent of the time, as compared to about 20 percent of the time for the RBC and ASP pilot plants during the first year of operation. The RBC and ASP pilot plants had the next lowest effluent SS concentrations. The TF/SC effluent SS was only less than 5 mg/L, for 5 percent of the time. Figure 3-34 shows that the effluent TP concentration was less than 0.5 and 1.0 mg/L for 96 and 99.5 percent of the time, respectively (during the first year).

During the follow-up extension period, nitrification improved when the media depth was increased to 2.5 m and additional phosphorus was added to the BAF starting on January 15, 1992. Figure 3-35 is a trend plot showing the effluent ammonia-N concentration during the follow-up extension period. After January 15, 1992 the effluent ammonia-N ranged from 1.0 to 5.0 mg/L.

The influent rates for organic and ammonia loading varied considerably due to variations in the influent concentrations and intentional changes in the flow rate. Flow rate changes occurred primarily from January 11 to March 11, 1991 when the flow was reduced to encourage nitrification in the unit. Figure 3-36 shows the actual and smoothed organic loading. The smoothed curve shows that the organic loading increased after October 29, 1990 because polymer was taken off the full-scale plant. After January 11, 1991 and until March 11, 1991, the loading reduced since the flow

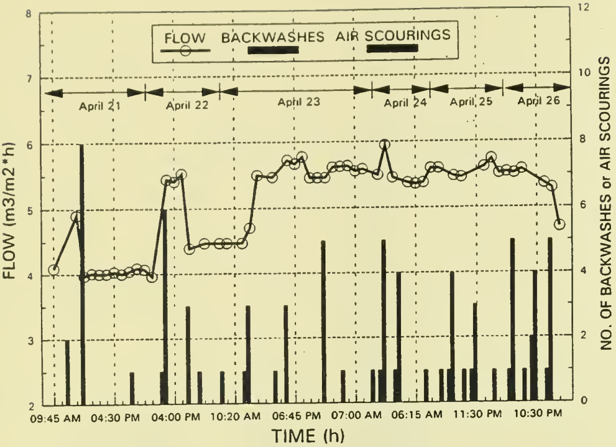


FIGURE 3.31 BAF HYDRAULIC TEST RESULTS

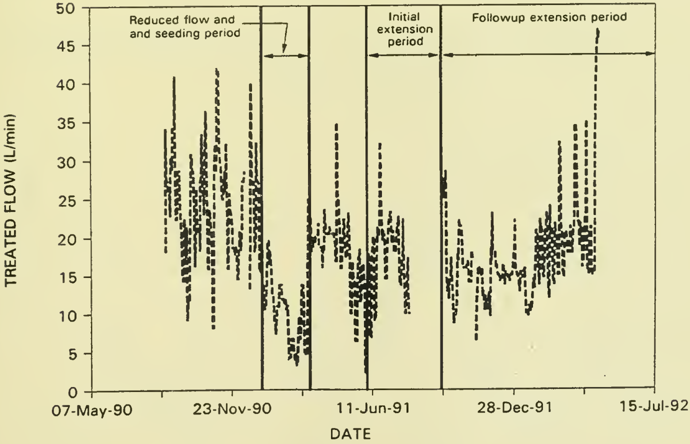


FIGURE 3.32 BAF AVERAGE DAILY TREATED FLOW

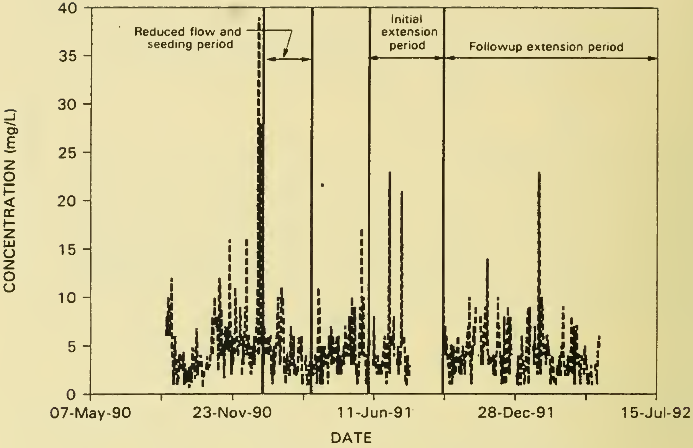


FIGURE 3.33 BAF FINAL EFFLUENT SS CONCENTRATION

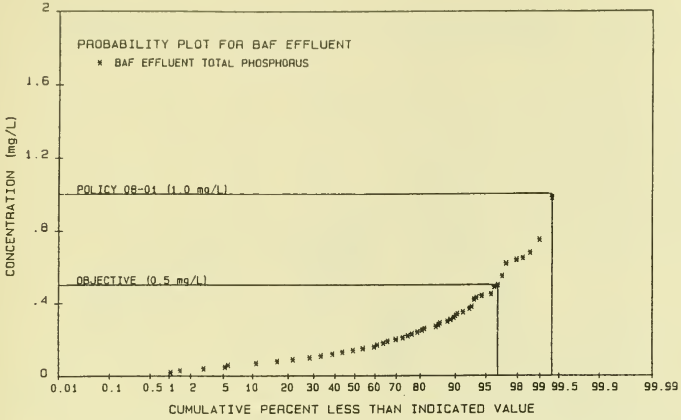


FIGURE 3.34 PROBABILITY PLOT BAF EFFLUENT TP CONCENTRATION

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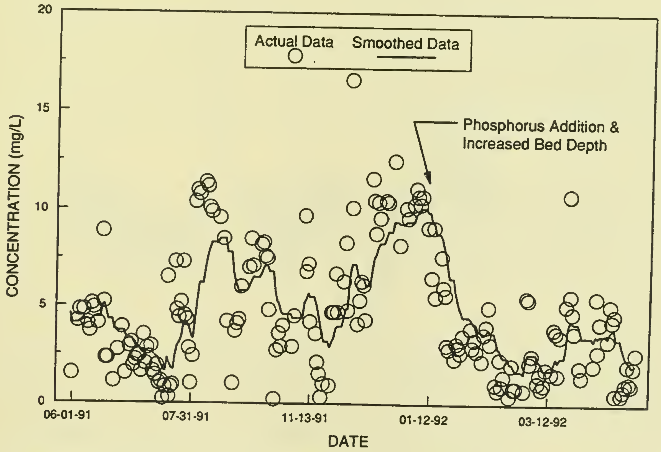

FIGURE 3.35 TREND PLOT OF BAF EFFLUENT AMMONIA-N CONCENTRATION DURING FOLLOW-UP EXTENSION PERIOD

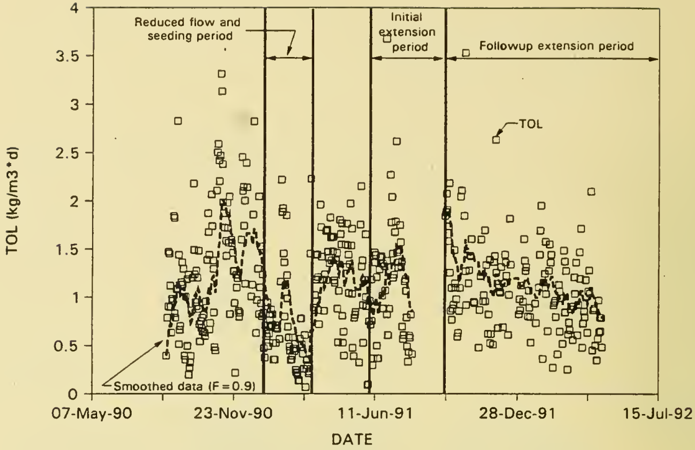

FIGURE 3.36 BAF TOTAL ORGANIC LOADING RATE

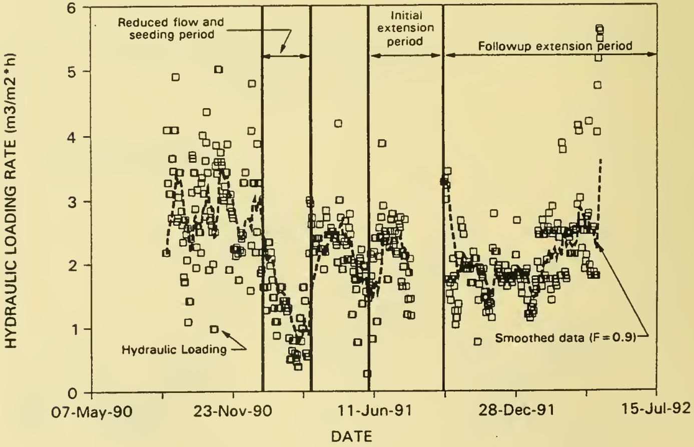

FIGURE 3.37 BAF HYDRAULIC LOADING RATE

was intentionally reduced to assist in re-establishing nitrification. After March 11, 1991, the loading was increased to an average of 1.2 kg $BOD_5/m^3 \cdot d$. During the follow-up extension period the TOL reduced to an average of 1.1 kg $BOD_5/m^3 \bullet d$.

Figure 3-37 shows the actual and smoothed hydraulic loading rate from August 1990 to April 1992. The plot shows that the hydraulic loading was initially slightly higher than the target averaging 2.8 $m^3/m^2 \bullet h$. The rate was reduced from January 11 to March 11, 1991 to as low as 0.5 $m^3/m^2 \bullet h$ to assist in reestablishing nitrification in the pilot unit. After March 11, 1991, the loading was maintained relatively constant around 2.4 $m^3/m^2 \bullet h$ until the unit was initially shut-down at the beginning of August 1991. During the follow-up extension period the hydraulic loading averaged 2.1 $m^3/m^2 \bullet h$, although this was increased to 2.4 $m^3/m^2 \bullet h$ after the media depth was increased to 2.5 m.

Nitrification Profiles

NH_3-N removal was investigated when the unit was nitrifying, through the fall of 1990, the spring of 1991, and the spring of 1992 by sampling the six ports that were spaced evenly along the BAF's depth. Figure 3-38 shows the results based on the retention time through the unit on each particular day. The curve represents a best-fit line for the pooled data during each period. The curves indicate that ammonia removal is constant through the unit. The rate appears to depend on the influent concentration, since effluent concentrations were similar. Nitrification rates were more rapid during July 1991 than January and February 1992 at a reduced media depth and at higher wastewater temperatures.

Relationship Between Loadings and Effluent Responses

Figure 3-39 is a trend plot of the organic loading and the effluent total BOD_5 concentration for the period August 1990 to May 1991. Figure 3-40 shows the organic loading versus the effluent total BOD_5 concentration and the relationship between the results; data are scattered. Temperatures did not indicate a relationship during the period. No relationship was evident between the organic loading and the NH_3-N removal during the last year of operation, possibly due to problems establishing nitrification.

Figure 41 shows the relationship between the ammonia loading and effluent ammonia-N concentration for the BAF, after the media depth was increased to 2.5 m. The figure shows a marginal relationship ($r^2 = 0.55$) indicating an increasing effluent ammonia-N concentration at an increasing influent nitrogen loading.

Solids Yield

Wasting from the BAF was accomplished by performing generally one backwash per day. Each backwash consisted of five or six air and water cleaning cycles. Effluent water was used for this purpose. The waste solids in the backwash water were low in concentration, averaging 349 mg/L. Figure 3-42 is a cumulative plot of the solids

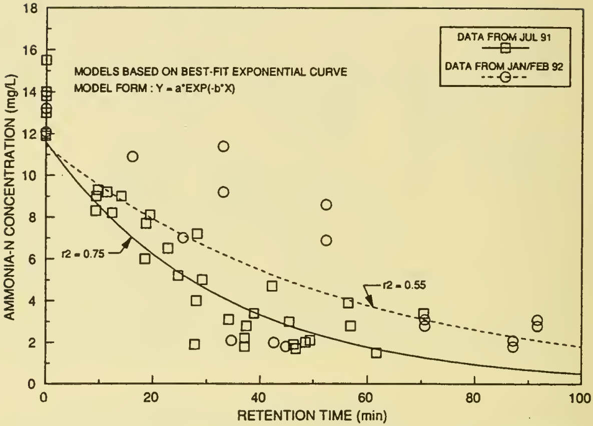

FIGURE 3.38 AMMONIA-N CONCENTRATION PROFILES THROUGH BAF TANK

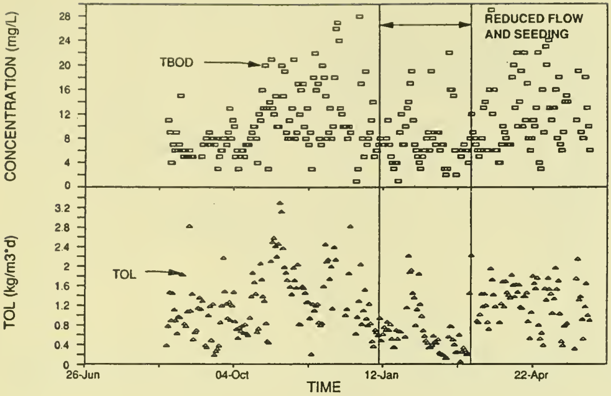

FIGURE 3.39 TREND PLOT OF ORGANIC LOADING AND EFFLUENT TOTAL BOD_5 CONCENTRATION

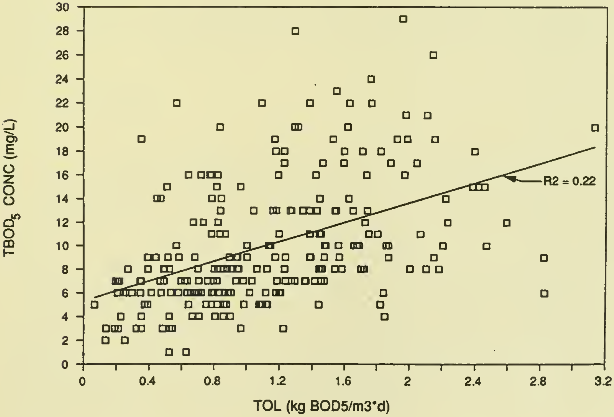

FIGURE 3.40 BAF ORGANIC LOADING RATE VERSUS EFFLUENT TOTAL BOD_5 CONCENTRATION

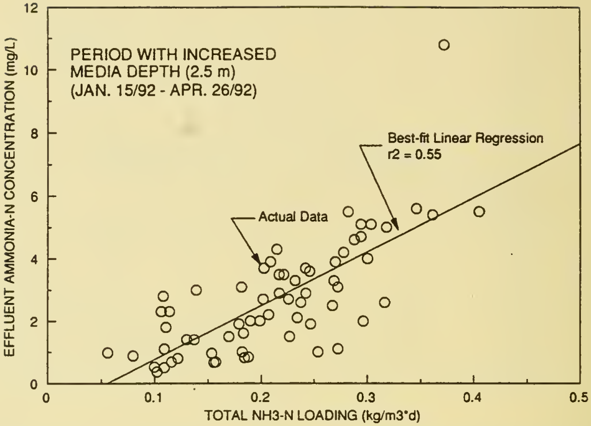

FIGURE 3.41 RELATIONSHIP BETWEEN INFLUENT AMMONIA-N LOADING AND EFFLUENT CONCENTRATIONS

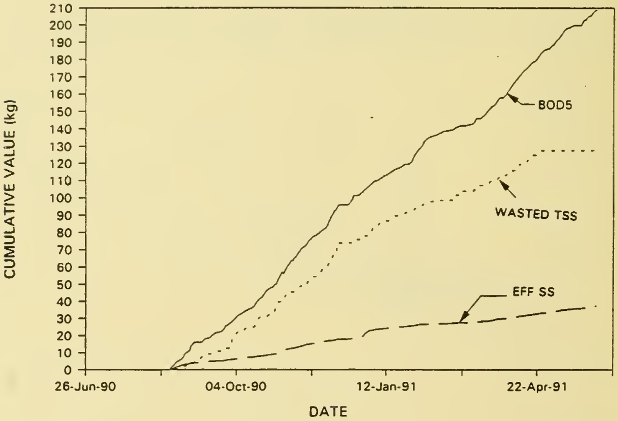

FIGURE 3.42 CUMULATIVE SOLIDS WASTED AND TOTAL BOD_5 REMOVED FOR BAF PROCESS

intentionally wasted, solids lost, and total BOD_5 removed from the unit for the period beginning August 20, 1990. The solids generated are the solids backwashed from the filter and wasted. The solids lost from the system are the solids unintentionally lost in the final effluent. The total BOD_5 removed is the difference between the influent and effluent total BOD_5 loading. The figure shows that about four times the amount of solids are wasted intentionally from the system than solids that are lost from the system in the effluent. Over this period, the overall solids yield was 0.79 kg TSS/kg BOD_5 removed, or 0.61 kg VSS/kg BOD_5 removed. Since this unit has a solids yield similar to the other units, the volume of the waste would be expected to be greater per unit of flow treated. For the test period, the BAF backwash averaged 2.1 L/min (0.25 $m^3/m^2 \bullet h$) or 11 percent of the treated flow.

ASP PROCESS

General

The ASP process was started at the end of July 1990; however, process mechanical problems persisted until mid-November 1990. Settleability in the final clarifier was a problem since it caused solids washout on a number of occasions. Improved final clarification was obtained by installing a flexible slow speed mixer in the bottom of the final clarifier on November 16, 1990. These mechanical problems were associated with the bench-scale size of the process and not the ASP. In addition, the initial HRT of 2.5 h which was used based on experience from the City's Little River PCP was increased to 3.5 h on October 30, 1991. The unit operated well until December 24, 1990 when the feed lines froze. The unit was restarted on January 11, 1991 with winter protection after which the unit operated continuously until it was shut down early in June 1991. Table 3.15 summarizes the influent parameters (wastewater flow, pH, and temperature) and the effluent concentrations for four periods corresponding to the overall period and the three periods of varying chemical and polymer usage in the full-scale plant clarifiers.

Hydraulic Loading

It is difficult to determine if the ASP was affected when polymer was removed from the two clarifiers feeding the pilot plants because the pilot plant had not been operating well before this point due to the operational mechanical problems mentioned above. The flow rate was reduced at the end of October 1990 to provide a 3.7 h HRT for aeration. On a diurnal flow basis, the ASP treated down to a 2.3 h aeration HRT on a daily basis. This is only a 1.5 diurnal peak factor, and resulted because this unit was not operational until the early winter. The unit did hydraulically handle the flow to a 1.7 h aeration HRT on October 19; however, the effluent concentrations on this day exceeded the objectives. The unit also exceeded the effluent objectives on normal flow days during this period. The Little River PCP has been shown to be able to handle a 2.0 peaking factor; therefore, it was felt that this unit would also handle peak flows. The pilot plant did treat the simplified diurnal flow pattern modelled after the full-scale plant on a daily basis.

Table 3.15
Summary of Performance of the ASP Pilot Plant

Period	Influent Parameters			Effluent Concentrations (mg/L)					
	Flow (mL/min)	pH	Temp (°C)	SS	Total BOD_5	Carbonaceous BOD_5	TP	NH_3-N	Un-ionized NH_3-N
Overall Period (Nov. 16/90 - May 31/91)	49	7.4	12	12	13	7	0.3	0.9	0.008
Alum + Polymer Addition (Aug. 8/90 - Oct 29/90)	77	7.2	17	39	29	17	1.0	1.2	0.006
Alum Without Polymer Addition (Oct. 30/90 to Jan. 10/91)	47	7.5	12	16	18	10	0.5	0.2	0.002
Ferric and Polymer Addition (Jan. 11/91 - May 31/91)	49	7.4	13	12	12	6	0.3	1.0	0.010

Effluent Quality

For the overall period, the ASP met the effluent objectives of 15, 15, 0.5, and 0.1 mg/L for SS, total BOD_5, TP, and un-ionized NH_3-N. The ASP also met the target of 2 mg/L total NH_3-N for this same period. The effluent NH_3-N concentration averaged 0.9 mg/L and was generally less than 2.0 mg/L. For comparison, the Little River PCP operated at effluent NH_3-N concentrations of less than 1.0 mg/L from November 1990 to March 1991[15].

Table 3.16 summarizes the ASP loading parameters. On average, the aeration tank loadings were close to the target loading rates, averaging an organic loading of 0.32 g $BOD_5/m^3 \cdot d$, an HRT of 3.9 h, and an F/M ratio of 0.22 d^{-1}. The SRT averaged 12 d from November 16, 1990 to May 31, 1991, higher than the target of between 7 and 9 d. This average resulted from the batch wasting operation.

The effluent concentrations were less than the effluent objective concentrations on a daily basis for 75, 68, 81, and 100 percent of the time for SS, total BOD_5, TP and un-ionized NH_3-N from November 16, 1990 to May 31, 1991. The ASP effluent met the effluent total NH_3-N target for 90 percent of the time. In terms of the effluent ammonia-N target of 3 mg/L, used during the follow-up extension period, the ASP was less than this NH_3-N target for 95 percent of the time. These comparisons with the effluent objectives are for comparative purposes only, since these refer to monthly average design objectives. In terms of an upper limit value, the effluent concentrations from the ASP were compared to Policy 08-01 values for SS, total BOD_5, and TP which are 25, 25, and 1.0 mg/L, respectively. Based on these limits, the ASP effluent was less than these limits for 92, 95, and 97 percent of time.

Trends

Figure 3-43 shows the flow treated by the ASP pilot plant. The apparent cyclic flow is due to weekend operation at the lower diurnal flow setpoint. Initially, the pilot plant was operated at about 70 mL/min (2.5 h aeration HRT), but this rate was reduced to 47 mL/min in November 1990. By the beginning of April 1991, the flow rate was increased to approximately 60 mL/min (2.9 h aeration HRT). Figure 3-44 shows the aeration tank HRT with the HRT starting at 2.5 h, increasing to 4.0 h and decreasing to 3.5 h near the end of the study. Therefore, the critical aeration tank HRT range is between 3.8 and 2.9 h plus approximately 1/3 to 1/2 h HRT in the preceding anoxic chamber.

The effluent quality from the ASP in terms of SS, total and carbonaceous BOD_5, and TP improved after November 16, 1990, due to operational changes. Figure 3-45 shows a trend plot of the effluent SS concentration, which reduced from an average of about 40 mg/L to 12 mg/L after November 16, 1990. Figure 3-46 shows the effluent carbonaceous BOD_5 concentration trend plot. The plot shows that effluent concentrations after November 16, 1990 averaged 7 mg/L. Figure 3-47 shows that for 90 percent of the time from November 16, 1990 to May 31, 1991, the ASP was nitrifying to less than 2 mg/L total NH_3-N.

Table 3.16
Summary of ASP Operating and Monitoring Parameters

Period	Operating Parameters					Clarifier		Monitoring Parameters	
	Organic Loading (kg $BOD_5/m^3 \cdot d$)	Anoxic HRT (min)	Aeration HRT (h)	F/M Ratio (d^{-1})	SRT (d)	SOR ($m^3/m^2 \cdot d$)	SLR (kg $SS/m^2 \cdot d$)	DO Concentration	MLSS (mg/L)
Overall Period (Nov. 16/90 - May 31/91)	0.32	31	3.9	0.22	12.3	4.1	12	7.4	2,139
Alum plus Polymer Addition (Aug. 8/90 - Oct. 29.90)	0.32	18	2.3	0.23	4.1	5.7	15	5.8	1,891
Alum Without Polymer Addition (Oct. 30/90 - Jan. 10/91)	0.32	31	3.9	0.13	12.8	3.1	15	7.0	2,838
Ferric and Polymer Addition (Jan. 11/91 - May 31/91)	0.32	31	3.8	0.25	11.9	4.1	11	7.5	1,916

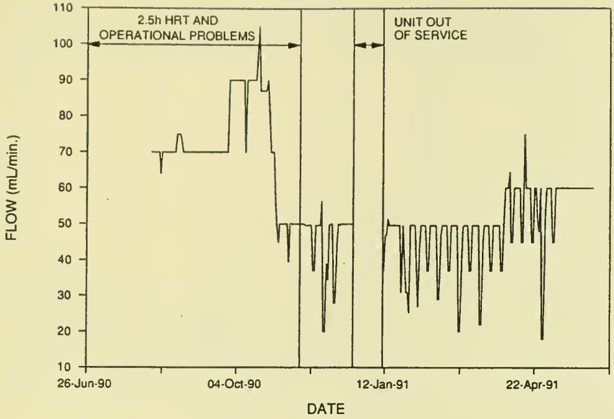

FIGURE 3.43 ASP AVERAGE INFLUENT FLOW

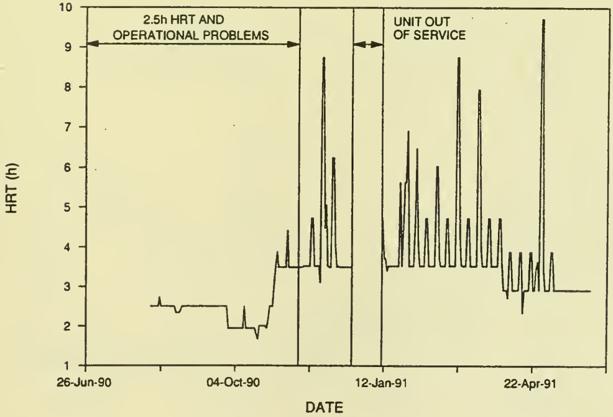

FIGURE 3.44 ASP AERATION TANK HYDRAULIC RETENTION TIME

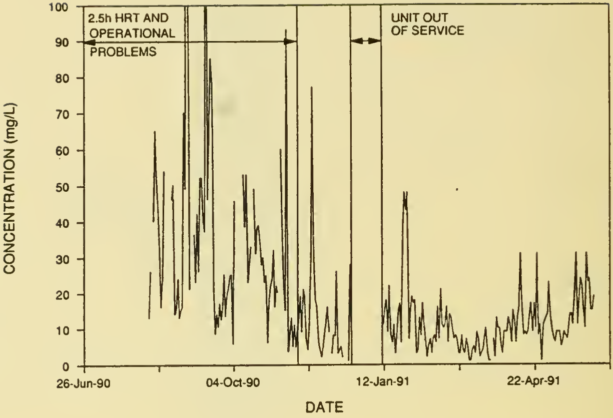

FIGURE 3.45 ASP EFFLUENT SUSPENDED SOLIDS CONCENTRATION

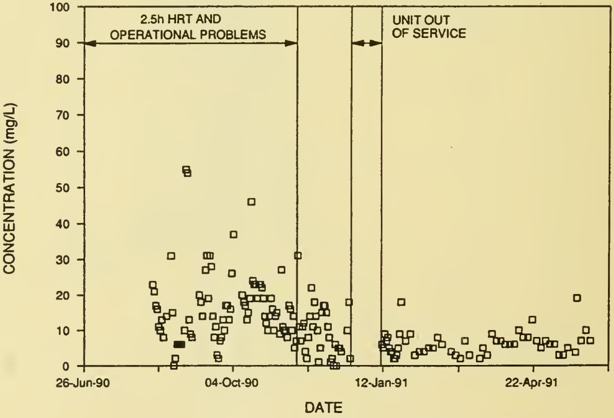

FIGURE 3.46 ASP EFFLUENT CARBONACEOUS BOD$_5$ CONCENTRATION

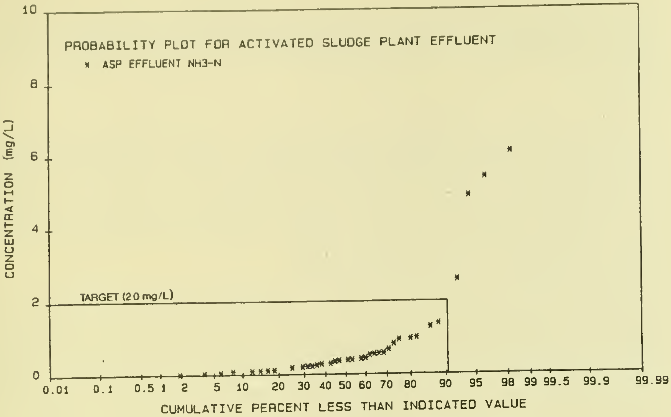

FIGURE 3.47 PROBABILITY PLOT OF ASP EFFLUENT AMMONIA-N CONCENTRATION

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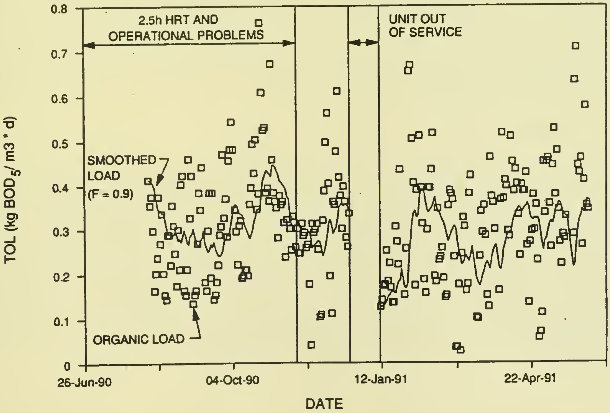


FIGURE 3.48 ASP INFLUENT ORGANIC LOADING RATE

The organic loading was near the initial target value of 0.34 kg $BOD_5/m^3 \cdot d$. The organic loading rate varied from approximately 0.1 to 0.8 kg $BOD_5/m^3 \cdot d$ on a daily basis. Figure 3-48 shows the trend plot of the organic loading for the entire operating period.

The F/M ratio was slightly higher than the initial target of 0.18 d^{-1}, averaging 0.22 d^{-1} over the period from November 16, 1990 to May 31, 1991. The F/M ratio dropped to an average of 0.13 d^{-1} from October 30, 1990 to January 10, 1991 when polymer was shut off from the full-scale plant. The reduced F/M ratio was due to higher MLSS concentrations during this period. Figure 3-49 shows the F/M ratio for the study period. Figure 3-50 shows that the MLSS concentration increased to an average of 2840 mg/L during the period of lower F/Ms.

The SRT ranged from 4 to 12 d on a daily basis. Figure 3-51 shows the daily calculated SRT and the cumulative calculated SRT for the entire period and from January 11, 1991. The SRT dropped to an average of 4 d until after November 16, 1990 because of the poor settling effluent and resulting high effluent SS concentration (averaging 39 mg/L). Once the flow rate was reduced to provide an average HRT of 3.8 - 3.9 h and the operational difficulties were corrected, the SRT averaged around 12 d. At this SRT, excellent nitrification was obtained. Excellent nitrification was also being obtained at the City of Windsor's Little River PCP[15]. From November 14, 1990 to March 12, 1990, plant No. 1 at the Little River PCP averaged an effluent NH_3-N concentration of 0.5 mg/L at a SRT of 7 d.

Process Loadings

The impact of various loadings and operating parameters on the effluent quality from the ASP were investigated; however, no apparent relationships were evident except for a correlation between the F/M ratio and the effluent NH_3-N concentration. Other influent parameters such as organic, hydraulic, and nitrogen loadings did not correlate well with total BOD_5, SS and NH_3-N.

Solids Yield

Wasting from the ASP was manually accomplished by wasting a portion of the MLSS. Figure 3-52 shows the cumulative solids generated, solids lost and total BOD_5 removed from August 20, 1990. The generated solids are the wasted solids from the aeration tank. The solids lost from the system are the solids unintentionally lost in the effluent. The BOD_5 removed is the difference between the influent and effluent total BOD_5 loading. The figure shows that about twice the amount of solids are wasted intentionally from the system as are lost from the system in the effluent. Over this period, the overall solids yield was 0.79 kg TSS/kg BOD_5 removed, or 0.61 kg VSS/kg BOD_5 removed.

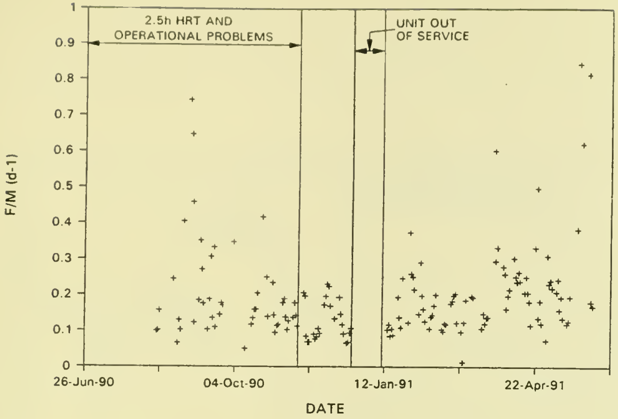


FIGURE 3.49 ASP FOOD-TO-MICROORGANISM RATIO

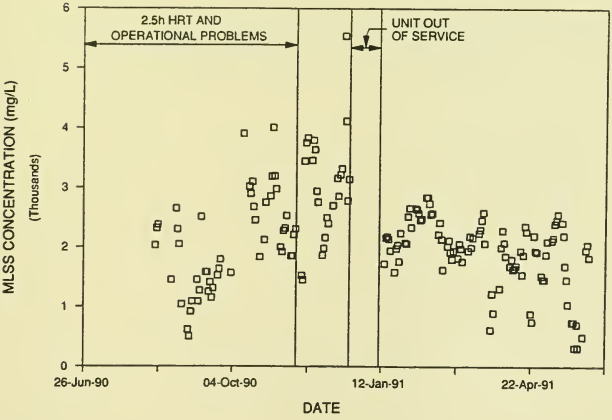


FIGURE 3.50 ASP AERATION TANK MLSS CONCENTRATION

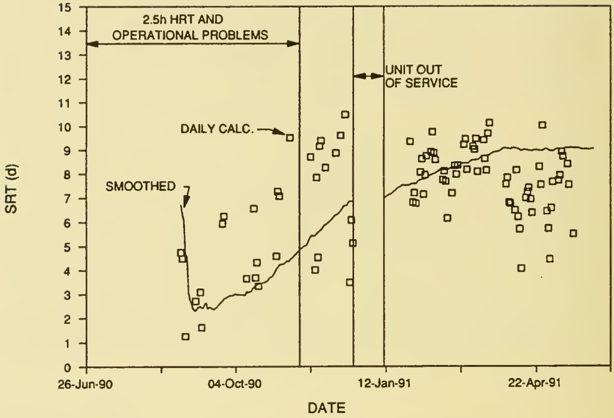


FIGURE 3.51 ASP SOLIDS RETENTION TIME

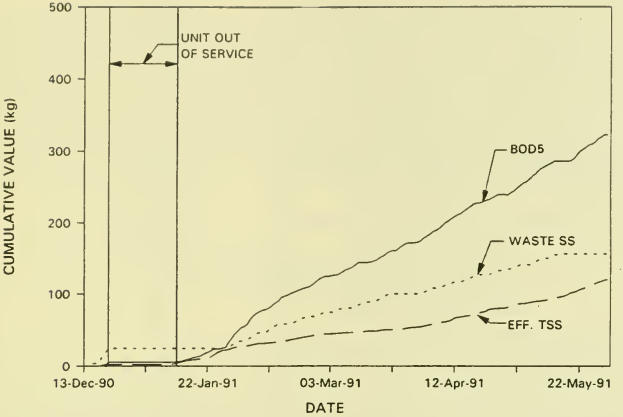


FIGURE 3.52 CUMMULATIVE SOLIDS WASTED AND TOTAL BOD_5 REMOVED FOR ASP PROCESS

PILOT PLANT SOLIDS HANDLING TESTS

The objective of the pilot plant solids handling tests was to evaluate the settling ability, thickening, and dewaterability of the biological sludge generated by each of the pilot plants.

The physical characteristics of sludges are the most important factors relating to sludge dewaterability. The basic physical parameters include specific gravity, solids concentration, water content, particle size, rheology and zeta potential. Because of the complex nature of sludges, these basic parameters are of only limited value in developing sludge design and operational data. Some other parameters have been developed to provide useful design information, particularly on the relative dewaterability of sludges. Among these are settleability, specific resistance, compressibility and Capillary Suction Time.

This section has been divided into three sections. The sections are:

- Bench-Top Settling Test Results
- Bench-Top Dewaterability Test Results
- Dewaterability testing

BENCH-TOP SETTLING TESTS RESULTS

Sludge settleability is a sludge characteristic mainly related to the physical properties, density, size and shape of the sludge solids. Knowledge of sludge settleability is used mainly in preselecting appropriate thickening processes, in operating activated sludge plants and determining the potential for chemical and polymer addition. In this section jar tests were used to observe the settling characteristics of each process's sludge.

The settling ability of raw sewage suspended solids, each biological sludge, and each biological sludge mixed with raw primary sludge was observed by conducting a series of jar tests. The tests were qualitative as described below. Samples with varying doses of chemical (ferric chloride) and anionic polymer (Secodyne 450w) were tested to determine the effect of conditioning agents on the sludge settling ability. The full-scale plant currently uses anionic polymer at dosages between 0.3 and 0.6 mg/L in conjunction with a precipitant for enhanced primary treatment. Settling tests were also conducted using two medium to high molecular weight cationic polymers (Secodyne 777 and Jeschem 2302). These tests indicated that a cationic polymer did not assist in settling the biological sludges. The dosages used for each series of settling tests is shown in Table 3.17. A series of settling tests were conducted using six parallel jars; the first jar was a control with no precipitant or polymer. The five remaining jars had varying dosages of precipitant and/or polymer.

The addition of either ferric chloride ($FeCl_3$) or polymer alone had no significant effect on the settling ability of any of the sludges; however, qualitative test data indicate that

Table 3.17
Chemical and Polymer Dosages Used for Setting Tests

No.	Sample	Dosage (mg/L)	
		Iron	Polymer
1	TF/SC	4	0.8
2	TF/SC	0	1.0
3	TF/SC	4	1.0
4	TF/SC	8	0.0
5	TF/SC	0	0.2
6	TF/SC	4	0.4
7	TF/SC	14	0.0
8	TF/SC	14	0.0
9	TF/SC	0	0.2
10	TF/SC	12	0.6
11	TF/SC	14	0.6
12	RAW+TF/SC	8	0.0
13	RAW+TF/SC	0	0.6
14	RAW+TF/SC	4	0.6
15	RAW+TF/SC	8	0.8
16	RAW+TF/SC	4	0.6
17	RAW+TF/SC	12	0.0
18	RAW+TF/SC	0	0.6
19	RAW+TF/SC	0	0.6
20	RAW+TF/SC	12	0.6
21	RBC	4	0.4
22	RBC	0	1.0
23	RBC	8	0.0
24	RBC	8	0.0
25	RBC	0	0.4
26	RBC	12	0.2
27	RBC	8	0.6
28	RBC	12	0.6
29	RAW+RBC	8	0.8
30	RAW+RBC	4	0.6
31	RAW+RBC	4	0.4
32	RAW+RBC	8	0.0
33	RAW+RBC	14	0.0
34	RAW+RBC	0	0.4
35	RAW+RBC	12	0.6
36	RAW+RBC	14	0.6
37	BAF	4	0.8
38	BAF	0	0.4
39	BAF	4	0.0
40	BAF	16	1.0
41	BAF	16	0.0
42	BAF	0	0.4
43	BAF	12	1.0
44	BAF	16	1.0
45	RAW+BAF	0	0.4
46	RAW+BAF	4	0.8
47	RAW+BAF	6	0.6
48	RAW+BAF	6	0.0
49	RAW+BAF	16	0.0
50	RAW+BAF	12	0.8
51	RAW+BAF	16	0.8
52	RAW+BAF	16	0.0
53	RAW+BAF	0	0.4
54	RAW+BAF	12	1.0
55	RAW+BAF	14	0.6
56	RAW	0	1.0
57	RAW	4	0.0
58	RAW	4	0.4
59	RAW	12	1.0
60	RAW	12	0.0
61	RAW	0	0.4
62	RAW	12	1.0
63	RAW	12	1.0
64	RAW	0	0.4

floc formation, settling characteristics, and turbidity improved significantly when both $FeCl_3$ and polymer were added. Best settling resulted from the addition of approximately 12 mg/L Fe^{+3} and 0.2 to 0.6 mg/L anionic polymer (Secodyne 450w). These dosages are similar to those used in the full-scale plant.

Although the BAF waste solids settled well, both on their own and with raw sludge added, the BAF waste (backwash) solids were investigated by conducting a series of pilot dissolved air flotation (DAF) tests. The DAF unit was considered because the BAF waste solids are much lower in concentration (average of 330 mg/L) than sludges from the other units. Bench-scale flotation testing is often used to predict the ability of thin sludges (< 1% solids) or sludges high in oil and grease content to thicken or dewater under dissolved air flotation. The bench-scale results give reasonable estimates of solids capture and subnatant quality and air to solids (A/S) ratios, however, float solids tend to be lower than obtained from full-scale systems. This can be attributed to the fact that the float is not removed continuously under bench-scale conditions. A series of tests were conducted at various chemical ($FeCl_3$) and polymer (anionic) dosages. The results are summarized in Table 3.18.

Table 3.18
Summary of DAF Testing for BAF Waste Solids

Test	Dosage (mg/L)		Float		Subnatant	
	Fe^{+3}	Secodyne 450w	Thickness (cm)	Solids (%)	Suspended Solids (mg/L)	BOD_5 (mg/L)
Control	0	0	0.4	2.64	32	32
Test 1	4	0.25	0.6	2.98	13	14
Test 2	8	0.50	0.7	3.32	12	NA
Test 3	16	1.0	0.7	3.86	11	NA

Note:
1 Backwash water initial suspended solids was 152 mg/L
2 Float solids results from the bench-scale units are usually lower than full-scale results. The floats on the bench-scale units were too thin to allow sampling without including subnatant.

The results show that a dosage of 4 mg/L Fe^{+3} and 0.25 mg/L anionic polymer produced a float or thickened sludge concentration of 3 percent and a subnatant SS concentration of 13 mg/L. Higher dosages of chemical and polymer do not improve the subnatant SS concentration significantly. Although a DAF unit could be used to thicken the waste solids from the BAF with good results, it is not essential.

Quantitative test data were inconclusive. Determination of dry solids in the sludge and suspended solids concentration in the subnatant/supernatant were difficult as separation was difficult. Excess supernatant in the solids distorted the reported sludge solids and carry-over of settled solids in the supernatant adversely affected the calculated suspended solids concentration.

Settled Volumes

The data were contradictory, generally indicating higher settling volumes for the control than for the best precipitant and polymer combination.

Sludge Solids

Data were scattered and generally showed higher sludge solids for the control sample than the treated sample. As indicated, sludge solids samples were difficult to obtain and this difficulty may have caused the problem. Also, the jar test apparatus provided only limited compaction compared to a full-scale clarifier.

Supernatant Solids

Data results were scattered, and generally lower suspended solids were measured in the supernatant of the treated samples than in the control. Jar tests were expected to show this result; but data may have been scattered due to resuspension of solids during sampling.

The settling ability of the TF/SC and BAF sludges improved when combined with raw primary sludge. RBC sludge co-thickened with raw primary sludge exhibited no significant change.

Overall, each biological sludge on its own or with raw sludge appeared to settle best and produce the best flocs when coagulant ($FeCl_3$) and polymer (anionic) were used. The optimal dosage was approximately 12 mg/L of Fe^{+3} and between 0.2 and 0.6 mg/L of anionic (Secodyne 450w) polymer. These dosages are similar to the dosages used in the full-scale plant.

BENCH-TOP DEWATERING TEST RESULTS

The relative ease with which water can be removed from a sludge by filtration provides a basis for designing sludge filtration equipment; for comparing the dewaterability of different sludges; and for evaluating the effect of various conditioning methods on a given sludge. The main parameters relating to sludge filterability are specific resistance, compressibility and Capillary Suction Time. A bench-top centrifuge test can also be used to evaluate the ability of sludges to be dewatered.

The dewatering characteristics of raw primary sludge, the biological sludge generated by the TF/SC, RBC, and BAF processes, and each biological sludge mixed with the raw primary sludge were determined by conducting a series of Capillary Suction Time (CST), Buchner Funnel, and bench-top centrifuge tests.

The CST test[18] provides a quantative indication of the filterability of a sludge. The CST test is a test that was developed to provide an alternative to the Buchner funnel test (for determining specific resistance) for measuring the filterability of sludges. The test consists of pouring a sample of sludge into the CST apparatus' reservoir which rests on air absorbent filter paper. The filtrate is drawn out of the reservoir by the capillary action of the paper. The liquid saturates the paper and spread outwards, in a circular pattern at a rate determined by the properties of the paper, the viscosity of the filtrate and the filterability of the sludge. The time taken for the liquid front to move a specified distance is measured with electrical probes and a timer. The CST test is simpler, less time consuming and requires less equipment than the Buchner funnel test. Since results can be obtained quickly and easily in the field it is very useful in selecting chemicals and dosage requirements and in process control in sludge conditioning systems.

Unfortunately, the CST test had to be discontinued because low suspended solids concentrations and free water in each sample made interpretation of the results inconclusive. Free water resulted because the thickened sludges were generated in small containers, not full-scale thickeners or clarifiers where sludge volumes would be more significant.

The bench-top centrifuge test[19] provides a qualitative indication of the relative effectiveness of separation by this method, but the results are not readily applicable to the actual performance during the operation of full-scale equipment. The test apparatus consists of two test tube size vessels into which conditioned sludge is measured. The samples are spun for a period of time and the supernatant and cake volume and concentration are determined.

The Buchner Funnel test[18] provided a quantitative measure (i.e. specific resistance) of the dewatering characteristic of a sludge when subjected to either pressure or vacuum filtration. In addition, the result of this bench-top test can be applied to the actual performance of full-scale operating equipment.

Specific resistance is defined as the pressure difference required to produce a unit rate of filtrate flow of unit viscosity through a unit weight of cake. The Buchner funnel test is used to determine values of specific resistance. The specific resistance value thus obtained provides a quantitative measure of the filterability of a sludge. A value of 1.0×10^{12} m/kg or less has been considered necessary for filtration to be an efficient method of dewatering. This should be used as a guideline only since many factors, including type of dewatering device (vacuum filter, belt filter or pressure filter) economics and availability of disposal alternatives will determine selection of dewatering equipment. Some typical specific resistance values for selected sludge are:

- Municipal primary ≈ 7 to 35×10^{13}

- Waste activated form P-removal system ≈ 2 x 10^{13}
- Partially digested primary and waste activated ≈ 7 x 10^{13}

Samples of TF/SC, RBC, and BAF sludge alone and mixed with raw primary sludge were tested with varying dosages of polymer to determine the effect of conditioning agents and co-thickening on sludge dewatering characteristics. The results of the Buchner Funnel and bench-top centrifuge tests are summarized in Table 3.19.

In all cases, the addition of polymer improved the dewaterability of the sludges. Quantitatively, this improvement is represented by lower values of specific resistance, higher cake solids, and lower suspended solids recorded in the centrate and higher volumes of centrate.

The test data suggested that the best dewatering was obtained by adding 1 to 2 kg of polymer per ton dry material (kg/tDM) to the sludge. The only exception to this trend was TF/SC sludge treated alone which dewatered best with 8 kg/tDM polymer addition. Although it was not an objective of the study to determine optimal polymer selection or doses, test data revealed that excess polymer addition was counter-productive, resulting in higher specific resistance values and increased centrate suspended solids concentrations.

DEWATERABILITY TESTING

Samples of biological sludge generated by the TF/SC and RBC processes were provided to Ashbrook Simon Hartley for dewaterability pilot plant testing. The RBC sludge was pretreated to solids concentrations representative of sludge from a full-scale RBC unit. The results are presented in Table 3.20.

The data from the pilot testing indicated that both the TF/SC and RBC sludges responded to gravity thickening and dewatering with conventional equipment. Bench-top tests indicated that solids settling and dewatering characteristics can be improved by combining raw primary sludge with the biological sludge generated by the TF/SC and RBC processes.

Table 3.19
Results of Bench-Top Settling Tests

Date	Sample Type	Pretreatment (mg/L)		Dewatering Polymer	Dosage (kg/tDM)	Specific Resistance (m/kg)	Centrate Suspended Solids (mg/L)	Cake Suspended Solids (%)	Volume Centrate (mL)
		Fe+3	450w Polymer						
22-May-91	BAF	16.0	1.0	None		3.9884E+13	136.1	5.52	36
22-May-91	BAF	16.0	1.0	None		5.7064E+12	52.63	20.15	38
22-May-91	BAF	16.0	1.0	None		5.2482E+13	85.7	5.73	35
22-May-91	BAF	16.0	1.0	Percol 728	2	9.0244E+11	63.2	6.81	38
22-May-91	BAF	16.0	1.0	Percol 728	4	8.3186E+11	69.2	6.99	39
22-May-91	BAF	16.0	1.0	Percol 728	8	1.5710E+13	105.1	7.1	39
22-May-91	BAF	16.0	1.0	Sec 750	1	1.0103E+12	56.4	6.99	39
22-May-91	BAF	16.0	1.0	Sec 750	2	N/A	63.4	7.79	41
22-May-91	BAF	16.0	1.0	Sec 750	2	1.8011E+11	13.95	6.42	43
22-May-91	BAF	16.0	1.0	Sec 750	2	1.1336E+12			
22-May-91	BAF	16.0	1.0	Sec 750	2	2.2040E+12	13.95	6.42	43
22-May-91	BAF	16.0	1.0	Sec 750	8	2.4860E+11	29.55	5.67	44
25-May-91	BAF & Primary (1)	16.0	1.0	None		8.6981E+12	302	9.8	44
25-May-91	BAF & Primary (1)	16.0	1.0	None		9.9408E+12	255	8.1	44
25-May-91	BAF & Primary (1)	16.0	1.0	Percol 728	2	7.3810E+11	52.3	10.6	44
25-May-91	BAF & Primary (1)	16.0	1.0	Percol 728	4	5.2927E+11	48.9	10.5	45
25-May-91	BAF & Primary (1)	16.0	1.0	Sec 750	1	6.0934E+11	24.4	11.5	45
25-May-91	BAF & Primary (1)	16.0	1.0	Sec 750	2	N/A	71.1	11.1	45
23-May-91	BAF & Raw	14.0	0.6	None		1.2222E+13	998	9.65	44
23-May-91	BAF & Raw	14.0	0.6	None		1.2640E+13	1278.2	9.36	43.5
23-May-91	BAF & Raw	14.0	0.6	Percol 728	2	4.8105E+11	53.49	10.88	43
23-May-91	BAF & Raw	14.0	0.6	Percol 728	4	8.8782E+12	88.63	10.72	44
23-May-91	BAF & Raw	14.0	0.6	Sec 750	1	8.0758E+11	54.9	10.63	45.5
23-May-91	BAF & Raw	14.0	0.6	Sec 750	2	N/A	70.5	10.47	44
21-May-91	RBC	8.0	0.6	None		1.4200E+13	189.3	5.88	37.5
21-May-91	RBC	8.0	0.6	None		1.7868E+13	113.5	5.83	37
21-May-91	RBC	8.0	0.6	Percol 728	2	5.0401E+11	76	6.79	39.5
21-May-91	RBC	8.0	0.6	Percol 728	8	7.4140E+13	102.3	6.82	43
21-May-91	RBC	8.0	0.6	Sec 750	2	3.5162E+13	55.4	7.07	41.5
21-May-91	RBC	8.0	0.6	Sec 750	8	9.5222E+13	132.5	6.59	40
24-May-91	RBC & Primary (2)	12.0	0.6	None		9.7998E+12	787	8.1	38.5
24-May-91	RBC & Primary (2)	0.0	0.0	None		1.1231E+13	510	8.6	39
24-May-91	RBC & Primary (2)	0.0	0.0	Percol 728	2	4.2939E+11	21.7	10	41.5
24-May-91	RBC & Primary (2)	0.0	0.0	Percol 728	4	8.7854E+11	57.8	9.45	45
24-May-91	RBC & Primary (2)	0.0	0.0	Sec 750	1	3.5938E+12	35.7	9.9	42
24-May-91	RBC & Primary (2)	0.0	0.0	Sec 750	2	N/A	59	9.8	44
23-May-91	RBC & Raw	12.0	0.6	None		6.1137E+12	780	8.02	40.5
23-May-91	RBC & Raw	12.0	0.6	None		5.5201E+12	809.8	9.28	41
23-May-91	RBC & Raw	12.0	0.6	Percol 728	1	9.8828E+11	35.7	10.11	42
23-May-91	RBC & Raw	12.0	0.6	Percol 728	2	9.7385E+12	20.9	10.22	43
23-May-91	RBC & Raw	12.0	0.6	Sec 750	1	6.0878E+11	25.9	9.69	42.5
23-May-91	RBC & Raw	12.0	0.6	Sec 750	2	N/A	57.8	9.74	39
10-Jul-91	TF/SC	12	0.6	None		3.2276E+12	10.3	6.02	39
10-Jul-91	TF/SC	12	0.6	None		3.0070E+12	28.2	6.93	39
10-Jul-91	TF/SC	12	0.6	Percol 728	2	1.1806E+12	5	6.59	40
10-Jul-91	TF/SC	12	0.6	Percol 728	4	1.7400E+12	4.8	7.37	42
10-Jul-91	TF/SC	12	0.6	Percol 728	8	4.3250E+11	23.3	6.72	43
10-Jul-91	TF/SC	12	0.6	Sec 750	2	2.6402E+12	2.5	6.76	40
10-Jul-91	TF/SC	12	0.6	Sec 750	4	1.6615E+12	28.2	6.66	41.5
10-Jul-91	TF/SC	12	0.6	Sec 750	8	8.3461E+11	47.6	6.55	42
11-Jul-91	TF/SC & Raw	12	0.6	None		5.1301E+12	89.2	10.3	41.5
11-Jul-91	TF/SC & Raw	12	0.6	None		5.3081E+12	134.1	9.4	41
11-Jul-91	TF/SC & Raw	12	0.6	Percol 728	2	3.5940E+11	65.1	10.7	43
11-Jul-91	TF/SC & Raw	12	0.6	Percol 728	4	1.0794E+13	90.9	10	44
11-Jul-91	TF/SC & Raw	12	0.6	Sec 750	2	6.8576E+11	55.8	10.9	43
11-Jul-91	TF/SC & Raw	12	0.6	Sec 750	4		70.5	10.3	44
10-Jul-91	TF/SC & Primary (3)			None		7.2701E+12	137.5	10.5	40
10-Jul-91	TF/SC & Primary (3)			None		7.8951E+12	128.2	10.4	39
10-Jul-91	TF/SC & Primary (3)			Percol 728	2	1.4346E+12	34.1	10.9	41
10-Jul-91	TF/SC & Primary (3)			Percol 728	4	1.3823E+13	61.4	10.4	44
10-Jul-91	TF/SC & Primary (3)			Sec 750	2	1.4413E+12	26.8	11.5	41
10-Jul-91	TF/SC & Primary (3)			Sec 750	4	N/A	93.3	10	45

Notes :
(1) Primary Pretreatment 12 mg/L Fe+3 and 1.0 mg/L 450W
(2) Primary Pretreatment 12 mg/L Fe+3 and 0.4 mg/L 450W
(3) Primary Pretreatment 12 mg/L Fe+3 and 0.4 mg/L 450W

Table 3.20
Summary of Manufacturer's Dewatering Tests

Sludge Type	Dewatering System	Feed Concentration (%)	Cake Dry Solids (%)	SS Capture (%)
TF/SC	Gravity thickener	0.93	5.5	95
RBC	Gravity thickener	0.38	6.5	99
TF/SC	Filter press	0.93	16-18	95
RBC	Filter press	0.38	13-16	95

Overall, the settling ability and dewatering tests showed that the solids generated by the pilot plant processes can be handled in a conventional manner. This includes gravity thickening on their own or co-settled and thickened with the raw sludge with coagulant and polymer conditioning. Dewatering of the thickened sludges also appeared to be feasible for the three processes tested. Dosages for both settling and dewatering are within typical ranges and did not vary significantly for each process tested.

OPERATION AND MAINTENANCE OF THE PILOT PLANTS

The following section discusses the operation and maintenance of each process, as these pertain to the pilot units. The regular operational parameters used to control each process are also expected to be required at full-scale. The maintenance requirements or problems associated with each process may or may not pertain to a full-scale facility. Items are identified if they are expected to be only applicable to the pilot facilities and not a full-scale plant.

TF/SC

The TF/SC wastewater temperature as well as the sludge blanket height in the final clarifier were monitored daily. The solids contactor monitoring included DO concentration, and a 30-minute settling test. Daily diurnal flow rates were determined for the TF, SC, and the SC return sludge. A visual inspection involved the condition of growth at the top and bottom of the TF and the appearance of the solids contactor "mixed liquor". In addition to the slime inspection at the top of the tower, the heads in the distribution system had to be hosed down (with primary effluent) to remove excessive slime build-up that could impede the distribution of primary effluent over the media.

The key operating parameters for the TF effluent were suspended solids concentration, BOD_5 removal, and nitrification. Ammonia profiles on the lower half of tower were constructed on a weekly basis beginning in the spring of 1991. The key operational parameters for the SC were dissolved oxygen (4 mg/L), MLSS concentration with a target of 2,000 mg/L, SVI with a target of <150 mL/g, SRT of 2 to 3 days, and wasting rate. The key parameters for the final clarifier were the sludge blanket height and clarity and the return sludge rates. Microscopic examinations of the mixed liquor were conducted periodically to determine the system's condition by observing the biological population. The quality of the final effluent at the surface of the clarifier was inspected on a daily basis.

An online dissolved oxygen meter with a chart recorder to monitor the solids contactor DO concentration would have determined whether or not the basin received excessive amounts of dissolved oxygen. An online meter is recommended for future pilot and/or full-scale work.

The population of filter flies was extensive at the top and around the base of the TF because ambient wind was absent, since the pilots were located in a roofed three-sided building (previously a compost drying building). In a full-scale system with the towers fully exposed to the weather and a superior distribution system to maintain a more evenly distributed wetting rate, filter flies probably would have been reduced.

A prolonged absence of primary effluent to the tower could result in large scale drying out of the growth on the media, resulting in a major sloughing when flow is restored to the tower. Recycle pumps should alleviate this concern.

RBC PROCESS

The RBC's temperature, DO concentration at the first and fourth stage, diurnal flow pattern and rates, and sludge levels were monitored daily. A visual inspection was also performed each day. The condition of the growth (slime) on the media, clarity at the surface of clarifier (i.e. floating sludge), and waste pump and feed pump condition were inspected visually. The shaft bearings were inspected visually for grease, excessive noise and temperature, and the gear oil drive reservoir was checked periodically. The shaft bearings were greased every six weeks.

The key parameters for operation were the appearance of the slime and the DO concentration.

During the pilot study, the only operational concerns were that a prolonged power failure could cause a weight imbalance due to irregular drying of the media once rotation is stopped. This would require the capability to drain the RBC tank below the lowest edge of the media to allow uniform drying prior to restarting to prevent media/shaft/gear failures.

BAF PROCESS

Daily monitoring of the BAF unit included temperature and DO concentration of the filtered effluent and/or adjustments to the process air flow rate and diurnal flow rates. In addition to the daily monitoring, periodic NH_3-N profiles consisted of taking a sample at each of the six sample points across the unit.

The system required backwashing each day; it was usually done each morning. The pilot plant normally required one backwash cycle per day; a cycle consisted of five backwashes. Periodically, when flow rates were elevated above normal (i.e. storm simulation) the pilot plant would require a mini-backwash cycle consisting of two or three cycles. The date, time, and number of backwashes were recorded. Each backwash required a volume of approximately 480 litres of filtered effluent which was stored in the 1,500 L backwash tank equipped with an overflow. During a backwash cycle, a composite sample of the backwash water, usually 20 litres, was collected and analyzed for TSS and TVS.

The key operating parameters for the pilot plant were the DO concentration in the final effluent and the liquid height above the media. When the DO concentration of the final effluent was below 1 mg/L, either the filter was clogging, resulting in poor water quality, or the process air flow rate was decreased. The liquid height above the media indicated when the filter was clogging or air bridging was occurring.

A level recorder which would indicate the liquid height above the media, monitor the filter clogging rate across the media, and detect when air bridging is occurring would have been useful, although foaming was sometimes a problem. A backwash sequence for daily cycle and mini-backwash would eliminate the need to initiate each of the backwash cycles manually. A feedback system for the process air and backwash air lines to maintain tighter control of air flow rates would also be beneficial.

The air bridging did not allow the pilot plant to operate to specification. The addition of an air scouring once a day appeared to correct the air bridging problem. Air scour assisted in breaking up the media and reducing clogging.

The air lines controlling the pneumatic valves during a backwash froze during the very cold days causing several problems. The electric solenoid valve on the backwash airline allowed water to back flow through the lines and flood the air rotometer and regulator. The air supply for the pilot plant was too humid which caused the regulators and rotometers to foul frequently. Media loss through backwashing (intentional and inadvertent) occurred and media had to be retrieved frequently from the drain channel. The process air at the bottom of the pilot plant seemed to channel up one side of the BAF on a regular basis. This channelling resulted in poor oxygen transfer and more frequent backwashing because the flow went through only part of the unit, although it is not expected that this will be a problem at full-scale.

ASP PROCESS

Due to the small size of the ASP pilot system, numerous hydraulic, freezing and settling problems had to be solved. The system was taken out-of-service a number of times for modifications that required reseeding. Since the unit only required ten litres of MLSS from Little River PCP and acclimation of MLSS to the primary effluent was very rapid, the down time was not significant. The problems that were encountered are unique to this bench unit only. Upon each startup, acclimation was rapid though the F/M ratio was below the target of 0.18 d^{-1} on a consistent basis.

Daily monitoring included daily diurnal flow patterns, aeration DO concentration, effluent temperature, and a 30-minute settling test on the mixed liquor. Key operating parameters were SRT and SVI.

SUMMARY AND COMPARISON OF PILOT PLANT PERFORMANCE

The preceding sections have summarized the operation of the pilot plants over the testing period; this section summarizes the performance of the pilot plants in order to evaluate and compare each process. Some of the processes were optimized further during the testing and/or had startup problems. In this section, performance is compared during the period when the process was optimized. This section is divided into the following main areas:

- Monthly Effluent Design Objectives
- Daily Effluent Performance compared to the Effluent Objectives and an Upper Limit
- Effluent Toxicity
- Hydraulic Performance
- Solids Handling
- Operational and Maintenance Observations

MONTHLY EFFLUENT DESIGN OBJECTIVES

The monthly effluent design concentrations or effluent objectives for the pilot plants were 15, 15, 0.5, and 0.1 mg/L for SS, total BOD_5, TP, and un-ionized NH_3-N, respectively. A target total NH_3-N concentration of 2 mg/L was set after the project commencement to estimate the actual degree of nitrification attained in each unit. During the follow-up extension period, the target ammonia-N concentration was reassessed at 3 mg/L. The monthly averages for these parameters and carbonaceous BOD_5 from August 1990 to April 1992 are shown in Table 3.21. The table shows that only the TF/SC and BAF processes were operated until the end of July 1991 and re-started and operated from September 1991 through April 1992. The RBC and ASP pilot plants were shut down at the end of May 1991. Only the RBC and BAF met the effluent design objectives for every monthly period for all effluent parameters; however,

Table 3.21
Pilot Plant Monthly Average Effluent Concentrations

Month	Concentration (mg/L)																							
	Suspended Solids				Total BOD5				Carbonaceous BOD5				Total Phosphorous				Un-Ionized NH3-N				Total NH3-N			
	TF/SC	RBC	BAF	ASP	TF/SC	RBC	BAF	ASP	TF/SC	RBC	BAF	ASP	TF/SC	RBC	BAF	ASP	TF/SC	RBC	BAF	ASP	TF/SC	RBC	BAF	ASP
Aug-90	16	7	7	32	21	11	8	23	10	6	5	12	0.5	0.3	0.1	0.8	0.013	0.019	0.029	0.008	1.2	1.9	3.9	0.9
Sep-90	14	7	3	51	16	11	7	30	9	6	6	18	0.6	0.4	0.1	1.3	0.019	0.014	0.018	0.014	1.8	1.8	2.0	2.0
Oct-90	13	8	4	30	18	14	10	33	10	8	5	20	0.5	0.3	0.1	0.9	0.020	0.013	0.007	0.003	1.9	1.9	1.1	0.5
Nov-90	28	9	6	21	26	14	15	20	16	8	8	12	1.1	0.4	0.4	0.7	0.009	0.007	0.042	0.002	1.5	1.4	6.1	0.3
Dec-90	16	8	7	8	19	14	13	13	9	8	10	8	0.6	0.4	0.3	0.3	0.005	0.011	0.042	0.002	0.8	2.2	6.0	0.1
Jan-91	16	7	5	18	13	10	8	15	6	5	6	6	0.6	0.3	0.2	0.5	0.005	0.009	0.057	0.003	0.9	2.7	8.9	0.4
Feb-91	14	6	4	10	15	9	8	11	6	5	6	5	0.4	0.2	0.1	0.2	0.009	0.013	0.043	0.004	1.8	3.8	9.7	0.5
Mar-91	13	6	3	6	17	8	8	9	7	5	7	6	0.4	0.2	0.1	0.1	0.007	0.009	0.059	0.006	1.2	2.3	9.1	0.8
Apr-91	18	11	4	12	17	12	12	15	10	8	7	8	0.5	0.3	0.1	0.4	0.006	0.009	0.047	0.017	1.1	2.1	6.7	1.8
May-91	20	7	5	14	24	11	13	11	12	6	8	7	0.5	0.3	0.1	0.4	0.021	0.014	0.045	0.014	2.2	2.3	6.1	1.5
Jun-91	14	-	4	-	14	-	10	-	7	-	5	-	0.4	-	0.1	-	0.016	-	0.033	-	1.4	-	3.6	-
Jul-91	15	-	5	-	18	-	11	-	7	-	8	-	1.0	-	0.7	-	0.013	-	0.019	-	1.2	-	2.8	-
Aug-91	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sep-91	12	-	5	-	22	-	11	-	-	-	-	-	0.3	-	0.1	-	-	-	-	-	-	-	-	-
Oct-91	18	-	4	-	26	-	9	-	13	-	6	-	1.1	-	0.1	-	0.018	-	0.035	-	2.8	-	5.2	-
Nov-91	16	-	5	-	26	-	8	-	15	-	7	-	0.4	-	0.2	-	0.024	-	0.023	-	3.8	-	4.1	-
Dec-91	16	-	4	-	25	-	10	-	14	-	4	-	0.3	-	0.2	-	0.017	-	0.035	-	4.4	-	8.7	-
Jan-92	20	-	5	-	38	-	6	-	12	-	-	-	0.3	-	1.3	-	0.038	-	0.024	-	8.2	-	6.8	-
Feb-92	20	-	4	-	28	-	9	-	17	-	4	-	0.5	-	0.5	-	0.042	-	0.006	-	7.2	-	2.3	-
Mar-92	15	-	4	-	21	-	10	-	8	-	5	-	0.5	-	0.2	-	0.009	-	0.010	-	2.7	-	3.2	-
Apr-92	16	-	3	-	24	-	7	-	15	-	4	-	0.4	-	0.1	-	0.005	-	0.007	-	2.7	-	2.5	-

notes : 1) * represents change made to TP average to remove high influent concentrations on 2 d. Actual effluent TP monthly averages were 1.0 and 0.7 for the TF/SC and BAF processes.
2) - indicates value not available since process not in operation during this month
3) period used for evaluation is shown in shade for each process, these periods are used in subsequent evaluation, ie shade is- ▨(example shading)

the target total NH_3-N concentrations were not met for these two processes for reasons described earlier.

As outlined previously, some of the pilot plants experienced startup problems for some periods; these periods are not used for evaluating each process. The TF/SC process was not operational as a combined process with the SC until December 18, 1990. In addition, a solids washout occurred at the beginning of May 1991, which resulted in almost a complete loss of solids from the SC tank. The SC tank was restarted in May 1991 and operated until the end of July 1991 and again from September 1991 to April 1992. During the follow-up extension period from September 1991 to April 1992, the TF/SC process did not perform as well as the previous period. The TF/SC did not perform as well because the focus during the follow-up extension period was in establishing a viable SC unit, not effluent performance. Therefore, the follow-up extension period is not considered in this evaluation.

The RBC started in July 1990 and operated continuously until the end of May 1991. The BAF ran continuously from mid-August 1990 to the end of July 1991 and again from September 1991 to April 1992. However, during the follow-up extension period, only the period after the media depth was increased has been evaluated. The hydraulic loading was significantly reduced from January 11, 1991 to March 11, 1991 to attempt to re-establish nitrification in the unit. The ASP had operational difficulties until mid-November 1990, and the unit was also out-of-service for two weeks between December 1990 and January 1991.

Table 3.22 outlines the percentage of time the pilot processes met the effluent objectives on a monthly basis, for the months when the processes were operating as planned or optimized. The table shows that the RBC and BAF processes met the required monthly effluent objectives for each design objective parameter for 100 percent of the time. The RBC and BAF met the effluent monthly objective for SS, total BOD_5, TP, and un-ionized NH_3-N for 100 percent of the time for the months evaluated. Both the RBC and BAF met the monthly target for total NH_3-N (2 mg/L) for 40 and 40 percent of the time, respectively. It should be emphasized, however, that total effluent NH_3-N was not a design objective but was a target objective for pilot-plant operational purposes. In terms of a 3 mg/L NH_3-N target, the RBC and BAF met this target 90 and 80 percent of the time, respectively.

The RBC was kept intentionally loaded at relatively high nitrogen loading throughout the winter months as partial nitrification continued to be attained. This provided more intensive data in connection with performance data for BOD_5, SS, and TP.

Similarly, apparent hydraulic bottlenecks in the pilot scale BAF unit resulted in its intensive backwashing which is believed to have resulted in the interruption of nitrification during the winter. During the follow-up extension, a target effluent ammonia-N concentration of 3 mg/L was considered adequate due to higher influent BOD_5 concentration. Once the media depth was increased to 2.5 m, the BAF effluent ammonia-N concentration was less than 3 mg/L for two of the three months tested. Both the ASP and the TF/SC process failed to meet the design objectives for BOD_5,

Table 3.22
Summary of Monthly Performance for Each Pilot Process

Process	Period	Percentage of Months within Monthly Design Objective						
		SS	Total BOD_5	Carbonaceous BOD_5**	TP	Un-ionized NH_3-N	Target NH_3-N*	
							2 mg/L	3 mg/L
TF/SC	6 Months January 1991 to April 1991 June 1991 to July 1991	66	50	100	83	100	100	100
RBC	10 Months August 1990 to May 1991	100	100	100	100	100	40	90
BAF	11 Months (8 months for NH_3-N) September 1990 to July 1991 (excluding January, February, March) February 1992 to April 1992	100	100	100	100	100	40	80
ASP	6 Months December 1990 to May 1991	83	83	100	83	100	100	100

Notes:

*Target ammonia-N concentration for BAF process based on performance prior to vigourous backwashing (Sept. 90 - Oct. 90) and with 2.5 m media depth in place.

**Comparison for carbonaceous BOD_5 made using 15 mg/L objective since no objective is available for this parameter.

SS, and TP on a consistent monthly basis even during the selected months when the units were considered optimized. The ASP met each effluent objective the majority of the time (i.e. greater than 80 percent). The TF/SC process met the process target total NH_3-N concentration of 2 mg/L, except during May 1991, but did not meet the monthly objective for SS and total BOD_5 the majority of the time. During the follow-up extension period the TF/SC performed worse than the initial period, since the focus was on attempts to establish a viable SC.

All processes had monthly average effluent carbonaceous BOD_5 concentrations of less than 15 mg/L, although no design objective was established for this parameter.

DAILY EFFLUENT PERFORMANCE

Daily effluent evaluation criteria has been included to show the spread of the effluent concentrations generated from the pilot plants with respect to the effluent design objectives and an upper limit concentration. Although the effluent design objective concentrations were set as monthly objectives, the percentage of time that the processes were able to produce a high quality effluent on a daily basis provides a measure of confidence or safety factor in terms of performance consistency. Confidence is also provided by determining the percentage of time a process produces an effluent less than an upper limit concentration. In this case, the MOE Policy 08-01 limits (secondary treatment limits) for SS, total BOD_5 and TP of 25, 25, and 1.0 mg/L were used.

The results of this daily comparative evaluation are summarized in Table 3.23. The table shows the percentage of time each pilot plant's effluent parameters were less than the objective and the upper limit concentrations. The periods of operation for each pilot plant are the same as discussed above (Table 3.21). The initial and follow-up extension period has been included for the BAF's extension period. The BAF nitrified for a greater percent of the time in the follow-up extension period, once the media depth was increased to 2.5 m. This improvement has been added to the evaluation. During the follow-up extension period, once the media depth was increased to 2.5 m, the BAF process nitrified consistently. The TF/SC was also operated during both extension periods, but its performance was similar or worse than the original operating period (December 1990 to May 1991), and therefore was not used.

The results show that the RBC and BAF performed consistently in terms of SS, total BOD_5, and TP. The RBC and BAF operated at less than the effluent objective concentrations for SS, total BOD_5, and TP more than 80 percent of the time and less than the upper limit greater than 98 percent of the time. Both of these processes met the un-ionized NH_3-N objective 100 percent of the time, but were within the NH_3-N objective of 2 mg/L for 55 and 53 percent of the time for reasons previously discussed. These two processes were less than a 3 mg/L NH_3-N objective for 74 and 71 percent of the time, respectively.

Table 3.23

Performance of the Pilot Plants on a Daily Basis

Process	Probability Less than (%)												
	Suspended Solids		Total BOD_5		Carbonaceous BOD_5		Total P		Un-ionized Ammonia-N		Target Ammonia-N		
	Objective	Upper Limit	Objective	Upper Limit	Objective	Upper Limit	Objective	Upper Limit	Objective	Upper Limit	Objective		Upper Limit
											2 mg/L	3 mg/L	
Concentration Target (mg/L)	15	25	15	25	15	25	0.5	1.0	0.1	-	2.0	3.0	--
TF/SC	55	92	51	89	97	100	64	98	100	-	82	92	--
RBC	96	99	81	99	98	100	92	98	100	-	55	74	--
BAF	98	99	85	98	96	98	94	>99	100	-	53*	71*	--
ASP	75	92	68	95	90	100	81	97	100	-	90	95	--

*Performance based on initial nitrification period (September 7, to October 31, 1990) and on increased media depth (January 22 to April 26, 1992)

The ASP process was within the effluent objective and upper limits concentrations for SS, total BOD_5 and TP greater than 75, 68, and 81 percent of the time, respectively. The ASP met the effluent objective for un-ionized NH_3-N 100 percent of the time and for total NH_3-N 90 percent of the time.

The TF/SC process was within the objective and upper limit concentrations for SS, total BOD_5 and TP for greater than 55, 51, and 89 percent of the time, respectively. The TF/SC met the objective for un-ionized NH_3-N 100 percent of the time and the process target for total NH_3-N 78 percent of the time.

The ASP and TF/SC processes met a 3 mg/L NH_3-N target 95 and 92 percent of the time, respectively.

It is clear from the effluent data that both the RBC and BAF achieved consistent effluent performance on a daily and monthly basis. The ASP was a bench-scale unit and improved secondary clarifier performance would be anticipated at full-scale; this would improve the ASP performance in terms of BOD_5 and TP in the particulate form as well as SS. The coupled TF/SC process performance appears to be limited by the efficient TF performance and resultant low loads to the SC unit. This results in particulate carry-over in the final effluent that did not appear to be improved by flocculent addition at pilot-scale.

EFFLUENT TOXICITY

At project initiation, effluent toxicity was not an effluent criteria, but a number of tests were conducted by the City on the pilot plant effluents through the study. The effluent from the RBC and BAF were sampled and toxicity tests for rainbow trout and Daphnia magna were conducted on three occasions; the TF/SC's effluent was sampled on four occasions. Since the ASP pilot plant did not produce enough effluent for the tests, the results of tests conducted on the effluent of the City's Little River PCP were used for comparison over the same period. Toxicity testing on the effluent of the Little River PCP were conducted on four occasions. The toxicity results are provided in Appendix G. Overall, the results indicated that the effluents were non-lethal and the LC_{50} was greater than 100 percent for each process for both rainbow trout and Daphnia magna on each occasion. It should also be noted that the primary effluent was also non-lethal.

HYDRAULIC PERFORMANCE

Each pilot plant's ability to handle hydraulic loadings was evaluated based on average day, peak day, and diurnal flows. In terms of average flow rates, each process appeared to be able to handle average design values which were initially determined for each process. The flow to the solids contactor and final clarifier of the TF/SC process was reduced through the study, but the hydraulic loading to the final clarifier was set ultimately at the expected SOR for secondary sedimentation. The hydraulic and organic loading to the SC was not reduced to the same degree because one of the original two tanks was taken out-of-service.

The RBC operated for the entire study period at its design hydraulic loading. The RBC's performance included a final clarifier SOR that was in excess of twice the usual requirement based on conventional secondary clarifier designs for suspended growth systems.

The BAF initially had problems handling the average design flow due to filter clogging. This problem was corrected by making operational changes to the system backwashing and was verified during the follow-up extension period.

The ASP was initially loaded at the limiting hydraulic rate determined at the City's Little River PCP. Subsequent flow reductions to the pilot plant showed the plant could operate well at a hydraulic rate greater than MOE guideline design values.

The testing showed that all the processes should be able to handle peak day flows of 2.2 times the average daily flow. The TF/SC and RBC had no difficulty handling the 2.2 peak factor during the fall of 1990 and the spring of 1991.

The BAF initially did have difficulty handling peak flows, but it was felt that these problems had been corrected by operational changes. As a process, the BAF, although it can handle peak day flows, may require a more conservative design and hydraulic safety factor to operate during sustained peak flow events. Hydraulic tests conducted during the follow-up extension indicated this process could handle a sustained 3 to 4 d event at a peak factor of 2.2.

The ASP pilot plant handled a peak factor of only 1.5 during the period that it was optimized. During the fall of 1990, the unit handled a peak flow of 2.0. However, during this period the ASP's effluent quality did not meet the effluent objectives for SS, total BOD_5, and TP. The City has operated its Little River PCP at peak flows up to the equivalent of a 2.2 peaking factor during a stressing period[16]; therefore, it is not expected that the ASP process at the West Windsor PCP would have difficulty handling peak day flows.

Each pilot plant handled the simplified diurnal flow pattern modelled after the flow treated by the full-scale plant. The TF/SC, RBC, and ASP pilot plants handled this pattern five days a week with no apparent difficulty. The BAF did have difficulty handling the diurnal variation, but it appears the difficulty was related to partial clogging of the pilot unit. This partial clogging was addressed by modifying the backwashing procedure for the last period of operation, from June to July 1991, when the problem appeared to have been corrected.

SOLIDS HANDLING

The solids yields and solids handling for each of the pilot plants was investigated during the testing period. The solids yields for each process were determined and these are summarized in Table 3.24 for the period August 1990 through May 1991. On average, the BAF and ASP produced 25 percent less solids per unit of total BOD_5 removed than the TF/SC and RBC. For an activated sludge plant, the yield generally varies from 0.4 to 0.8 kg VSS/kg BOD_5, averaging 0.6 kg VSS/kg BOD_5[20]. Studies[16,17] conducted at the Little River PCP, Plant No. 1, indicate yields of 0.6 to 0.8 kg VSS/kg BOD_5. Therefore, the solids yields from the pilot plants are within expected values for a biological system.

Table 3.24
Pilot Plant Solids Yields

	kg TSS/kg BOD_5	kg VSS/kg BOD_5
TF/SC	1.1	0.8
RBC	1.0	0.8
BAF	0.8	0.6
ASP	0.8	0.6

In terms of solids intentionally wasted and solids unintentionally lost, the BAF process performed better than the other processes. The BAF intentionally wasted 3.4 times the amount of solids unintentionally lost. The RBC, ASP, and TF/SC intentional waste was 2.5, 1.5, and 1 times the amount of solids unintentionally lost, respectively.

In terms of solids handling, bench-top settling and dewaterability tests were conducted for the TF/SC, RBC, and BAF. The ASP pilot plant did not produce enough sludge to be included in this testing. However, the Little River PCP appears to have no problems co-settling its primary and secondary sludge with dewatering of the combined sludge by centrifugation. The testing results indicated that none of the processes appeared to have problems related to conditioning, settling, and dewatering on their own or co-settled with raw sewage. Settled sludges appeared to thicken with polymer. These bench-top tests were useful for determining if the processes could have a potential problem with respect to solids settling or dewatering. Further testing will be required at full-scale to determine actual full-scale conditioning and performance.

OPERATIONAL AND MAINTENANCE OBSERVATIONS

Some difficulties associated with the operation of each of the pilot facilities could be attributed to their scale. However, other difficulties and operational requirements appear to have been caused by specific processes. This aspect of performance is addressed below.

The TF/SC was more complex than the RBC and ASP. The TF/SC was a coupled process combining fixed film (TF) and suspended growth process operation; therefore, the operation and control requirements of this system were greater than the RBC and ASP pilot plants' requirements. In addition, the operation of the SC component of the process was difficult given the low loading from the TF tower. This problem may be addressed by increasing TF loading or by reducing chemical usage in the full-scale facility. The coupled process did provide a measure of additional nitrification.

The operation and maintenance requirements for the RBC pilot plant were lower than the other three processes' requirements. Minor mechanical checks of the RBC were required and minimal monitoring was conducted. However, operations cannot be changed if the process does fail. The RBC process generally has no mechanism to increase or decrease growth or provide additional aeration. However, full-scale design could provide variable or multi-speed drives and supplemental aeration.

The BAF also required intensive operation and maintenance, although most problems were related to the pilot scale size and lack of automated backwashing. Full-scale facilities provide an automated backwash system. The system also had problems with media washing out of the system during backwashing; it is unknown whether this is a concern at full-scale.

The ASP pilot plant was difficult to operate, but the majority of its problems could be related to the bench-top size of the unit. Problems such as clarifier sludge withdrawal, broken lines, and freezing were functions of the pilot plant, not the process. Based on the City's experience at the Little River PCP, the ASP process can be controlled well, but requires an average amount of operational and maintenance attention.

In summary, it was felt that based on the pilot plant operations, the RBC process was the easiest to operate, maintain and control. The ASP process was rated second to operate, maintain, and control based on the City's experience at the Little River PCP. The TF/SC and BAF required more intensive O&M; however, they did not require an amount of operation and maintenance large enough to prevent them from being considered at the West Windsor PCP. Also site visits to full-scale TF/SC and BAF plants indicated that these processes are viable from an O&M perspective.

Section 4

FULL-SCALE FACILITY PROCESS AND COST IMPLICATIONS

The full-scale plant process implications have been addressed by summarizing the site inspections made to several full-scale facilities and by estimating the space requirements at full-scale of each process on the West Windsor site. The space requirements are based on the pilot plant results. Full-scale plant implications are also addressed by summarizing the capital and O&M costs expected for each process.

FULL-SCALE SITE INSPECTION SUMMARY FOR POTENTIAL PROCESSES

To obtain input on the processes at full-scale, a number of RBC and BAF facilities in Ontario and Quebec were inspected. TF/SC sites in Michigan, Ohio, Texas, Utah and Colorado were also inspected. No ASP facilities were inspected: this process is reviewed based on the full-scale performance of the City's Little River PCP. Site inspection reports are contained in Appendix H.

TF/SC

Eight TF/SC plants with hydraulic design capacities ranging from 5700 to 236,600 m^3/d, were inspected in the United States. All the plants were secondary or tertiary (effluent filtration) treatment facilities. Flow equalization was practised at two of the facilities. Effluent requirements varied with each facility. None of the plants added precipitant/coagulant or polymer upstream of the primary clarifiers. Table 4.1 summarizes seven of the plants with respect to their design hydraulic capacity, current flow rate, influent and effluent concentrations, and loadings based on the results of the inspections and discussions with the operations staff. Each plant was operating below its design hydraulic capacity.

The trickling filter influent total BOD_5 concentrations for the seven plants varied from 88 to 159 mg/L (20 percent removals have been assumed for plants that only reported raw concentrations). Although the TF influent total BOD_5 concentrations were higher at each of these plants than the primary effluent from the West Windsor PCP, the TOLs for these plants are similar to the TOLs used for the West Windsor pilot plant. The TOLs for the seven plants varied from 0.12 to 0.72 kg $BOD_5/m^3 \cdot d$; whereas, the TOL at the TF/SC at West Windsor averaged 0.38 kg $BOD_5/m^3 \cdot d$.

The TF wetting rates at two of the plants, Rowlett Creek and Central Valley, were similar to the 29 $m^3/m^2 \cdot d$ value used for the West Windsor pilot plant; however, the TOLs at these two sites were almost double the West Windsor TF loading. The other

Table 4.1
Summary of TF/SC Process Plant Tours

Parameter	Plants						
	Wauseon Ohio	Coldwater Michigan	Duck Creek Texas	Rowlett Creek Texas	Central Valley Utah	Salt Lake City Utah	Boulder Colorado
1) Design Capacity (m^3/d)	5678	10200	113600	49200	236563	211960	60600
2) Actual Current Flow (m^3/d)	3785	7900	82500	42000	166919	126987	49600
2a) Flow Equilization	NO	NO	YES	YES	NO	NO	NO
2b) Nitrification Mandate	--	--	YES	YES	NO	NO	PARTIAL (separate stage)
2c) Effluent Filtration	NO	YES	YES	YES	NO	NO	NO
3) Influent Concentration (mg/L)	PE	RAW	PE	PE	NA	RAW	RAW
– TSS	88	150	159	73	NA	65	126
– TBOD5	113	200	137	111	NA	110	135
– TP	NA	2.3	NA	NA	NA	NA	NA
– NH3–N	14.5	NA	17.7	15.1	22	NA	12.9
4) Effluent Concentration (mg/L)							
– TSS	9	6	5	5	9	16	22
– TBOD5	5	10	3	5	5	14	16
– TP	NA	0.2	NA	NA	NA	NA	NA
– NH3–N	1.5	NA	0.6	3.1	5	NA	10.1
5) Type of Media	PLASTIC	PLASTIC	PLASTIC	PLASTIC	PLASTIC	ROCK	PLASTIC/ ROCK
6) Loadings							
– TOL (kg BOD5/m^3*d)	0.12	0.18	0.43	0.72	0.64	0.48	0.24
– Wetting Rate (m^3/m^2*d)	4.61	2.73	14.42	23.79	26.39	NA	5.39
– SC HRT (min)	NA	NA	73	181	103	20	136
– SOR (m^3/m^2*d)							
– at avg Q	NA	NA	23.4	12.3	18.3	NA	18.7
– at peak Q (PF=2)	NA	NA	46.7	24.6	36.6	24.4	37.5

notes : 1) NA indicates not available

plants had significantly lower wetting rates compared to West Windsor, varying from 3 to 14 $m^3/m^2 \cdot d$.

The SC hydraulic retention times were higher than the values used at West Windsor, except at the Salt Lake City plant. The solids contactor HRT at the Salt Lake City plant averaged about 20 min.

The peak clarifier SORs for the seven plants varied from 24 to 47 $m^3/m^2 \cdot d$; however, all peak SORs were based on handling a peaking factor of 2. Duck Creek had the highest peak SOR with this assumption and this facility had influent flow equalization. Therefore, the actual peak factor for this plant is less than 2. Discussions with the staff at the Central Valley plant, indicated that the process can operate at higher SORs. Intentions are to expand this plant with only limited or no expansion of the secondary clarification capacity. This expansion is proceeding and the peak SOR is expected to be as high as 65 $m^3/m^2 \cdot d$.

In terms of operational flexibility and site layout, the following observations were noted. All of the TF/SC plants visited had TF influent pumping stations and had the ability to recycle TF effluent back to the TF tower. Recycle was used at most of the plants to maintain a constant wetting rate of the filter media during low flow periods. Effluent lift pumps were installed where effluent filtration was practised.

Some plants had covers over the TF towers, although a number of the plants in colder locations did not have covers. Specifically Wauseon, Central Valley, and Salt Lake City, did not have covers on the TF towers.

Most plants co-settled the secondary waste solids in the primary clarifiers or gravity thickened the waste biological solids in a separate thickener. The Central Valley plant used gravity belt thickening to thicken secondary waste solids prior to digestion. For sludge dewatering, three of the plants used filter presses, one used centrifuges, one used drying beds, and one did not dewater sludge.

The main operating problem identified by the operations staff of the plants visited pertaining to the TF/SC process was a snail problem. Problems result from snail shells that accumulate in the channels and tanks downstream of the TF causing headloss and air diffuser/tank volume problems. Snails grow in the TF environment and the shells settle in various locations in the system. Removal is difficult if not planned for in the plant's design. Some of the sites also experienced filter flies (psychoda), but most of the plants that experienced this problem had covered TF towers.

RBC

RBC plants were visited in Guelph and Niagara Falls. Neither plant has an influent loading similar to the West Windsor PCP. The Niagara Falls RBC plant provides secondary treatment without nitrification. The organic and hydraulic loadings for the

Niagara Falls plant are estimated at 10.5 g $BOD_5/m^2{\bullet}d$ and 169 $L/m^2{\bullet}d$. The design peak SOR is 26 $m^3/m^2{\bullet}d$. The Guelph plant has separate stage nitrifying RBCs which follow an ASP and precede final effluent filtration. At the Guelph plant, the NH_3-N and hydraulic loading rates are estimated at 2.1 g $TKN/m^2{\bullet}d$ and 104 $L/m^2{\bullet}d$. The average loadings for the West Windsor RBC pilot plant were 3.5 g $BOD_5/m^2{\bullet}d$, 1.2 g $TKN/m^2{\bullet}d$ and 84 $L/m^2{\bullet}d$. Therefore, both of these full-scale plants were at a higher loading for organic carbon or NH_3-N removal. The application at the West Windsor PCP requires a process that will remove organic carbon and provide nitrification in a single step process.

The inspected RBCs' operation and maintenance procedures were informative. The RBCs at the Niagara Falls plant were individually driven by electric motors (5 hp each); at the Guelph plant the RBC units were air driven. Both plants indicated that the RBC units are a low maintenance item that require minimal attention and only periodic greasing. The Niagara Falls RBC units have been in service for 3½ years. Neither RBC plant required internal pumping due to the minimal hydraulic requirements of the RBCs.

An RBC facility with similar loadings to the West Windsor PCP is located in New Jersey. The 29,000 m^3/d Seven Mile WWTP adds ferric chloride upstream of its primary clarifiers for odour control (H_2S precipitation) and obtains good removal of SS and total BOD_5. Table 4.2 summarizes the influent flow, influent and effluent concentrations and loading rates of this facility. The plant does not test for NH_3-N; therefore, it is unknown if the plant nitrifies. The plant is rated at 29,000 m^3/d hydraulically and treats a seasonal loading. The peak loading occurs during the summer tourist season. The raw sewage at 115 mg/L BOD_5 is more concentrated than West Windsor's sewage. The winter primary effluent total BOD_5 concentration is 50 to 100 percent greater at 60 mg/L than the Windsor pilot plant influent total BOD_5 concentration. The TOL averaging 3.4 g total $BOD_5/m^2{\bullet}d$ is similar to the RBC pilot plant operated. Since the influent concentrations are greater than at West Windsor, the hydraulic loading at design is correspondingly lower than that treated by the pilot plant, averaging 42 $L/m^2{\bullet}d$.

BAF

Three full-scale BAF or BioCarbone plants were visited in Quebec with rated hydraulic capacities of 125,000, 25,000 and 2,200 m^3/d. The largest facility is located at Sherbrooke where construction was finished in 1991. The Sherbrooke plant was visited twice, first during construction on May 8, 1990 and again on May 3, 1991 during plant startup. Both the 25,000 m^3/d Boisbriand and the 2,500 m^3/d Canton d'Orford plants have been in operation for 2 and 4 years, respectively.

Limited analytical data from the Sherbrooke and Boisbriand plants is summarized in Table 4.3. The Sherbrooke plant data is based on a 20-day period during startup trials. The Boisbriand data is for a 2½-month period. Neither plant is required to nitrify. Organic loading rates at these two plants is about twice the rate sustained by the pilot

Table 4.2
Performance Summary of Seven Mile WWTP

Parameter	Value
FLOWS (m^3/d)	
- Design	29,000
- Current Winter Daily Average*	9,464
- Current Summer Daily Average	22,712
RAW INFLUENT (mg/L)	
WINTER	
- TSS	169
- Total BOD_5	115
SUMMER	
- TSS	211
- Total BOD_5	160
PRIMARY EFFLUENT (mg/L)	
WINTER	
- TSS	60
- Total BOD_5	60
SUMMER	
- TSS	95
- Total BOD_5	95
FINAL EFFLUENT (mg/L)	
WINTER	
- TSS	7
- Total BOD_5	6
SUMMER	
- TSS	10
- Total BOD_5	7
RBC LOADINGS	
ORGANIC ($gBOD_5/m^2 \cdot d$)*	3.4
DESIGN HYDRAULIC LOADING ($L/m^2 \cdot d$)	42
AVERAGE CLARIFIER SOR ($m^3/m^2 \cdot d$)	
WINTER	7.1
SUMMER	17.0

* RBC loading based on discussion with plant superintendent, not calculated

plant, although both the Quebec plants were designed for organic removal only. The average hydraulic loading rates are similar to those used for the pilot plant. The pilot hydraulic loading rate averaged 2.5 $m^3/m^2 \cdot h$.

Table 4.3
Summary of Full-Scale BAF Data

Parameter	Plant	
	Sherbrooke	Boisbriand
1. Treated Flow (m^3/d)		
- Raw influent	50,489	13,584
- Backwash	2,540	2,129
- Total	53,029	15,723
- Backwash percent of effluent	5	16
2. Influent Concentrations (mg/L)		
- BOD_5	55	83
- TSS	30	57
3. Effluent Concentrations (mg/L)		
- BOD_5	7	9
- TSS	5	8
4. Loadings		
- Organic ($kgBOD_5/m^2 \cdot d$)	2.2	2.1
- Hydraulic ($m^3/m^2 \cdot h$)	3.3	1.9

The main liquid train of the Sherbrooke plant operates by gravity. Pumping is only required for backwashing the filters. The BAF cells at Sherbrooke were covered, but there appeared to be no critical process reason for covering the filter surfaces. Waste solids from backwashing are equalized in a storage tank, prior to being fed into the primary clarifiers for co-settling with the raw sewage. Raw sludge is thickened in a gravity thickener; dewatering is accomplished by a plate and frame filter press.

Additional data for full-scale operating BAF (Biocarbone) systems in Europe is available in the literature.

ASP

As indicated above, no specific inspections were made of an ASP, since the City operates an ASP at its Little River PCP. The Little River PCP is rated at 63,600 m^3/d and adds precipitant for phosphorus removal upstream of the primary clarifiers. This results in a primary effluent averaging 59 mg/L BOD_5, lower than most typical ASPs[17]. The Little River plant has been loaded at higher rates than the ASP pilot plant. During the period November 1990 to March 1991, one of the two plants at the Little River PCP was loaded at an organic loading of 0.45 kg $BOD_5/m^3 \cdot d$, an HRT of 3.5 h and an SRT of 7.1 d[17]. At these loadings, the plant maintained an average effluent of 5, 5, 0.3, and 0.5 mg/L for SS, total BOD_5, TP, and NH_3-N respectively.

The plant's waste activated sludge is pumped to the inner ring of the primary clarifiers and co-mixed and settled with the primary sludge. Combined raw and waste activated sludge is dewatered by centrifuges and lime stabilized; this process is similar to the solids handling process currently used at the West Windsor PCP.

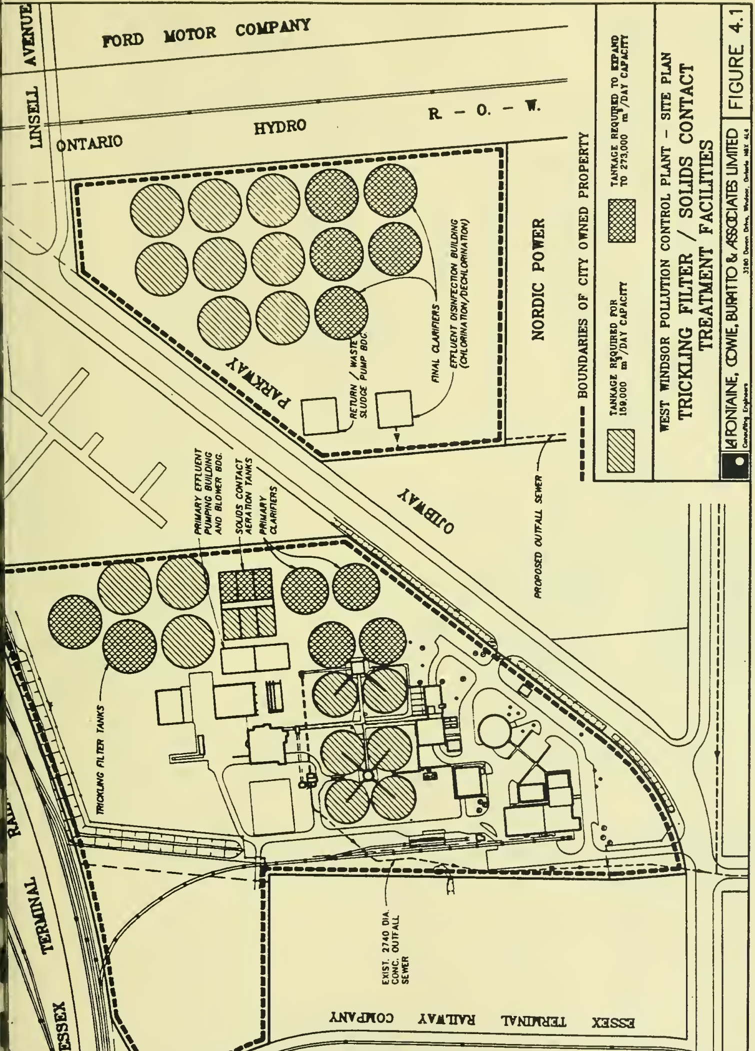
SITE AREA REQUIREMENTS BASED ON PILOT STUDIES

The preliminary site area requirements for each process were estimated based on the performance of the pilot plants. Process scale-up factors were not used. The site area requirements for comparison purposes are based on the plant's current hydraulic design capacity of 159,000 m^3/d and the ultimate site capacity of 273,000 m^3/d, both with a 2.2 peak factor. The preliminary sizing of each process based on the pilot plant testing is outlined in Table 4.4. Preliminary layouts of the upgrade at the West Windsor PCP are shown in Figures 4.1 to 4.4. Actual layouts should be determined at the plant pre-design stage. It must be noted that items such as road allowances and separation distances are not included. These items must be estimated at the pre-engineering stage. Hydraulic channels and piping arrangements must be decided upon in addition to headloss characteristics based on physical arrangements of the facility.

The data show that the smallest land area would be required by the BAF process, based on its performance during the pilot testing and resulting from treatment in one step without final clarification (Figure 4-3). The next largest facility, in terms of area required is an RBC plant (Figure 4-2). An RBC facility would have a significantly smaller site layout if the final clarifier requirements determined by the pilot testing are realized. Both the TF/SC and ASP layouts (Figure 4-1 and 4-4) indicate that these two processes would require approximately the same area which is governed largely by the secondary clarifier requirement and by intermediate and recycle pumping for a TF/SC. In terms of additional equipment, the TF/SC, RBC and ASP would be required to provide pumping of the existing primary effluent, after which the main liquid process would flow by gravity. Pumping is required since, to provide for the expected ultimate capacity, the final clarifiers will be located across the Objibway Parkway. The TF/SC process will require a larger pump capacity to provide enough head to pump to the top of the tower and provide subsequent gravity flow. The BAF process may be able to flow by gravity from the existing primary effluent. The existing plant was designed for an ASP expansion and for the combined facility to operate by gravity flow. Other processes may require initial pumping, if the soil conditions make it difficult to place the process tanks below grade. If site conditions require tanks to be elevated, the

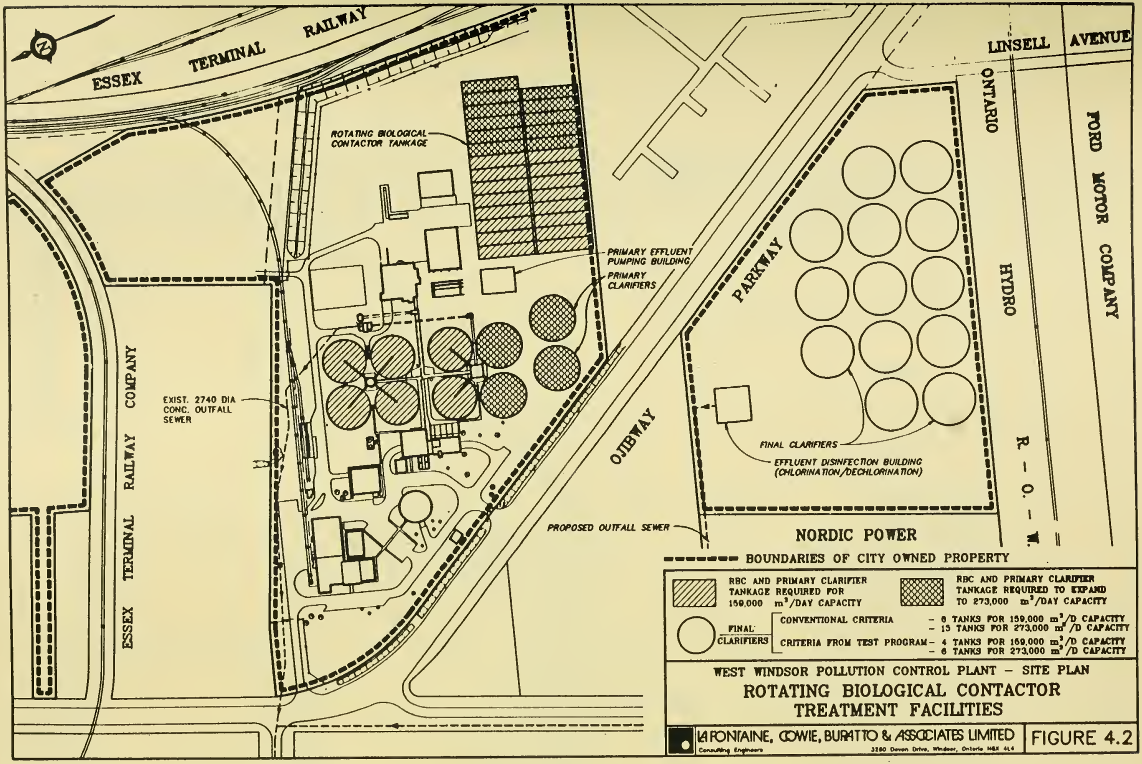
Table 4.4
Preliminary Full–Scale Plant Requirments Based on Pilot Plant Operations

Facility	Unit Process	Pilot Plant Results		Full–Scale Requirements for 159,000 m^3/d			Full–Scale Requirements for 273,000 m^3/d		
		Treated Flow (m^3/d)	Pilot Size	Full–Scale Size	Possible Configuration	Actual Size	Full–Scale Size	Possible Configuration	Actual Size
TF/SC	Trickling Filter	175	25.4 m^3	23078 m^3	– three TFs, each 45m dia and 5m tall	23856 m^3	39624 m^3	– five TFs, each 45m dia and 5m tall	39761 m^3
	Solids Contactor	60	2 m^3	5300 m^3	– three tanks, each 15m x 24m x 5m deep	5400 m^3	9100 m^3	– six tanks, each 15m x 24m x 5m deep	10800 m^3
	Final Clarifiers	60 (peak flow = 132 m^3/d)	4.5 m^2	11925 m^2	– eight tanks, each 45m dia	12723 m^2	20475 m^2	– thirteen tanks, each 45m dia	20676 m^2
RBC	RBC units	54	734 m^2	2154572 m^2	– 16 trains of 10 shafts per train and 13,973 m2 per shaft (high density)	2235680 m^2	3699360 m^2	– 27 trains of 10 shafts per train and 13,973 m2 per shaft (high density)	3772710 m^2
	Final Clarifiers	65 (peak flow = 143 m^3/d)	2 m^2	4892 m^2	– four tanks, each 45m dia	6362 m^2	8400 m^2	– six tanks, each 45m dia	9543 m^2
BAF	BAF units	30	0.5 m^2	2650 m^2	– 40 cells, each 84 m2 surface area and 2.5 m depth, additional area provided for : * backwash flow * cleaning	3360 m^2	4550 m^2	– 68 cells, each 84 m2 surface area and 2.5 m depth, additional area provided for : * backwash flow * cleaning	5712 m^2
ASP	Selector	0.072	0.0014 m^3	3092 m^3	– three tanks each, 10m x 25m x 5m deep	3750 m^3	5308 m^3	– five tanks each, 10m x 25m x 5m deep	6250 m^3
	Aeration Tanks	0.072	0.0104 m^3	22967 m^3	– six tanks, each 10m x 80m x 5m deep	24000 m^3	39433 m^3	– ten tanks, each 10m x 80m x 5m deep	40000 m^3
	Final Clarifiers	0.072 (peak flow = 0.0159 m^3/d & SORpeak = 29.4 m^3/m^2*d)	0.0054 m^2	11925 m^2	– eight tanks, each 45m dia	12723 m^2	20475 m^2	– thirteen tanks, each 45m dia	20676 m^2


FORD MOTOR COMPANY
LINSELL AVENUE
ONTARIO HYDRO R. – O. – W.
NORDIC POWER
PARKWAY
OJIBWAY
FINAL CLARIFIERS
EFFLUENT DISINFECTION BUILDING (CHLORINATION/DECHLORINATION)
RETURN / WASTE SLUDGE PUMP BDG.
PROPOSED OUTFALL SEWER
PRIMARY EFFLUENT PUMPING BUILDING AND BLOWER BDG.
SOLIDS CONTACT AERATION TANKS
PRIMARY CLARIFIERS
TRICKLING FILTER TANKS
EXIST. 2740 DIA. CONC. OUTFALL SEWER
ESSEX TERMINAL RAILWAY COMPANY
ESSEX TERMINAL RAIL
N
BOUNDARIES OF CITY OWNED PROPERTY
▨ TANKAGE REQUIRED FOR 150,000 m³/DAY CAPACITY
▩ TANKAGE REQUIRED TO EXPAND TO 273,000 m³/DAY CAPACITY
LAFONTAINE, COWIE, BURATTO & ASSOCIATES LIMITED
Consulting Engineers
3280 Devon Drive, Windsor, Ontario N8X 4L4

WEST WINDSOR POLLUTION CONTROL PLANT – SITE PLAN
TRICKLING FILTER / SOLIDS CONTACT TREATMENT FACILITIES

FIGURE 4.1


ESSEX TERMINAL RAILWAY
N
LINSELL AVENUE
ONTARIO HYDRO R. – O. – W.
FORD MOTOR COMPANY
ROTATING BIOLOGICAL CONTACTOR TANKAGE
PRIMARY EFFLUENT PUMPING BUILDING
PRIMARY CLARIFIERS
PARKWAY
OJIBWAY
FINAL CLARIFIERS
EFFLUENT DISINFECTION BUILDING (CHLORINATION/DECHLORINATION)
EXIST. 2740 DIA CONC. OUTFALL SEWER
ESSEX TERMINAL RAILWAY COMPANY
PROPOSED OUTFALL SEWER
NORDIC POWER
BOUNDARIES OF CITY OWNED PROPERTY
RBC AND PRIMARY CLARIFIER TANKAGE REQUIRED FOR 159,000 m³/DAY CAPACITY
RBC AND PRIMARY CLARIFIER TANKAGE REQUIRED TO EXPAND TO 273,000 m³/DAY CAPACITY
FINAL CLARIFIERS
CONVENTIONAL CRITERIA – 6 TANKS FOR 159,000 m³/D CAPACITY
– 13 TANKS FOR 273,000 m³/D CAPACITY
CRITERIA FROM TEST PROGRAM – 4 TANKS FOR 159,000 m³/D CAPACITY
– 6 TANKS FOR 273,000 m³/D CAPACITY
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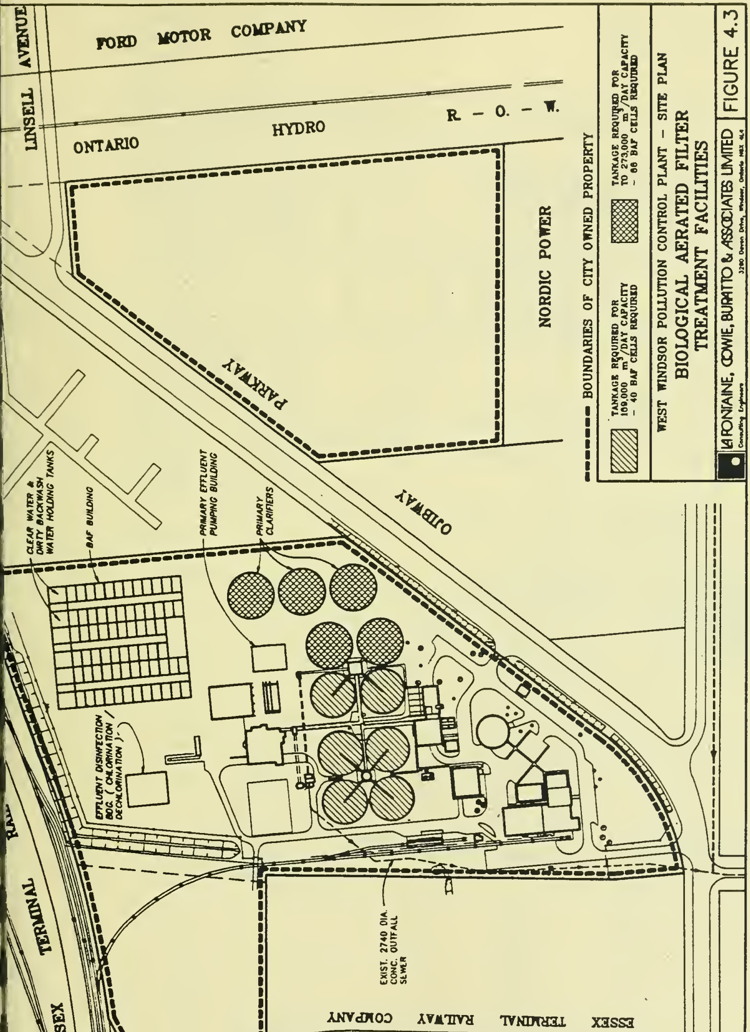
Consulting Engineers
3260 Devon Drive, Windsor, Ontario N8X 4L4

WEST WINDSOR POLLUTION CONTROL PLANT – SITE PLAN

ROTATING BIOLOGICAL CONTACTOR TREATMENT FACILITIES

FIGURE 4.2


FORD MOTOR COMPANY
LINSELL AVENUE
ONTARIO HYDRO R.-O.-W.
NORDIC POWER
OJIBWAY PARKWAY
CLEAR WATER & DIRTY BACKWASH WATER HOLDING TANKS
BAF BUILDING
PRIMARY EFFLUENT PUMPING BUILDING
PRIMARY CLARIFIERS
EFFLUENT DISINFECTION BDG. (CHLORINATION / DECHLORINATION)
EXIST. 2740 DIA. CONC. OUTFALL SEWER
ESSEX TERMINAL RAILWAY COMPANY
ESSEX TERMINAL RAIL
N
BOUNDARIES OF CITY OWNED PROPERTY
▨ TANKAGE REQUIRED FOR 109,000 m³/DAY CAPACITY - 40 BAF CELLS REQUIRED
▩ TANKAGE REQUIRED FOR TO 273,000 m³/DAY CAPACITY - 66 BAF CELLS REQUIRED

WEST WINDSOR POLLUTION CONTROL PLANT – SITE PLAN

BIOLOGICAL AERATED FILTER TREATMENT FACILITIES

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Consulting Engineers
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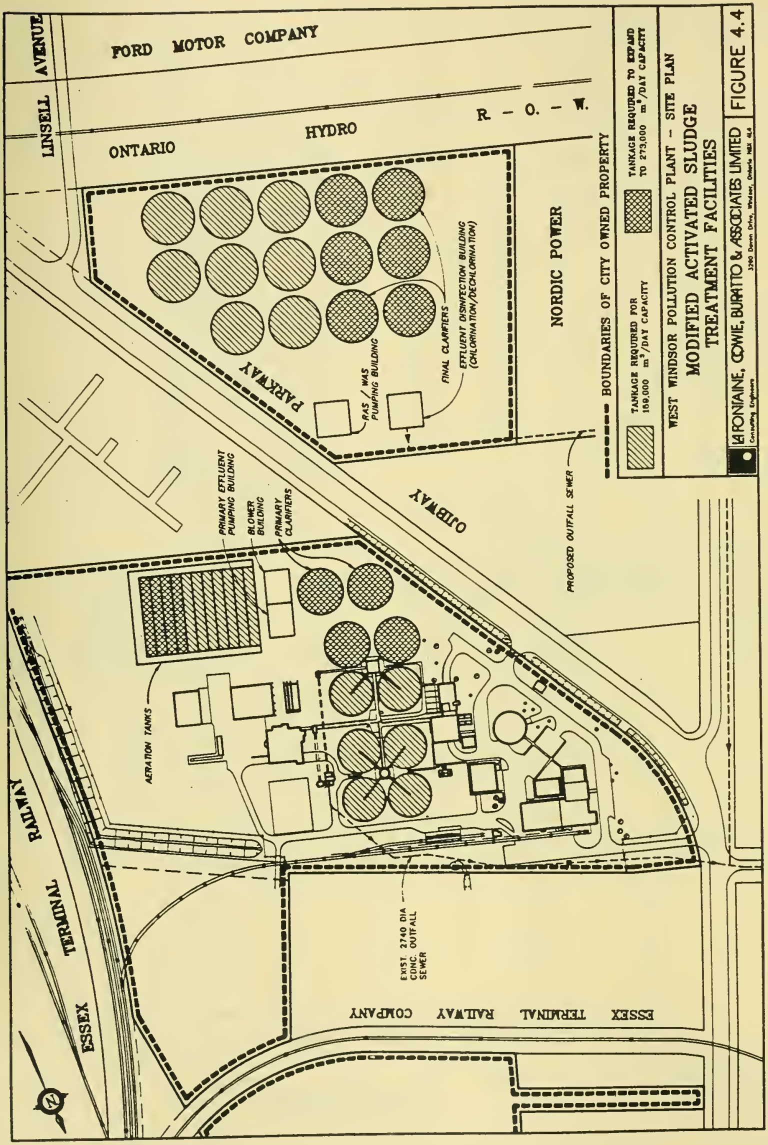
FIGURE 4.3


LINSELL AVENUE
FORD MOTOR COMPANY
ONTARIO HYDRO R. – O. – W.
NORDIC POWER
PARKWAY
OJIBWAY
ESSEX TERMINAL RAILWAY
ESSEX TERMINAL RAILWAY COMPANY
RAS / WAS PUMPING BUILDING
FINAL CLARIFIERS
EFFLUENT DISINFECTION BUILDING (CHLORINATION/DECHLORINATION)
PROPOSED OUTFALL SEWER
PRIMARY EFFLUENT PUMPING BUILDING
BLOWER BUILDING
PRIMARY CLARIFIERS
AERATION TANKS
EXIST. 2740 DIA CONC. OUTFALL SEWER
N
BOUNDARIES OF CITY OWNED PROPERTY
▨ TANKAGE REQUIRED FOR 169,000 m³/DAY CAPACITY
▦ TANKAGE REQUIRED TO EXPAND TO 273,000 m³/DAY CAPACITY
LaFONTAINE, COWIE, BURATTO & ASSOCIATES LIMITED
Consulting Engineers
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WEST WINDSOR POLLUTION CONTROL PLANT – SITE PLAN
MODIFIED ACTIVATED SLUDGE TREATMENT FACILITIES

FIGURE 4.4

TF/SC plant may require additional pumping after the TF tower, or the TF tower may have to be elevated an additional 2 to 3 m. Except for the BAF, a future requirement to provide for tertiary filtration would also require additional pumping.

The BAF process could be located on the existing site, although the advantages, except for a smaller plant size, have not been evaluated in this study.

It is also recommended that the City verify future sludge handling requirements to determine the need for onsite space needs, including existing building requirements.

COST IMPLICATIONS OF EACH PROCESS AT FULL-SCALE

Full-scale cost estimates for each process tested have been generated by LaFontaine, Cowie, Buratto and Associates Ltd. The cost estimate report is provided in Appendix J and is summarized in the following section. Cost estimates for capital and operating and maintenance (O&M) costs have been generated for comparison purposes and do not include processes or tasks common to all processes.

The cost estimate report indicates the following conditions and qualifications must be taken into consideration:

- Facilities common to all alternatives are not included in the cost estimates. Examples of common elements include: chlorination - dechlorination facilities; site work including access roads, drainage and process sewers; upgrading and expansion of the existing plant electrical distribution and instrumentation and control systems; potable and process water facilities.
- No allowance is included in the costs for known poor soil conditions at the West Windsor site. Cost estimates for the activated sludge alternatives are based on Little River Pollution Control plant construction costs where the soil bearing capacity is 3,000 lbs/ft^2 or greater. The soil bearing capacities at the West Windsor site are more typically 2,000 lbs/ft.^{2} or less and vary with depth and location. Special provisions are necessary at the West Windsor site to accommodate differential settlement.
- No costs are included for standby electric power generator facilities.
- No allowances are made for any changes in sludge handling, digestion or dewatering.
- The BAF sizing information and layout shows spare cells to allow for cell cleaning. The spare cells are not included in the cost estimate since other alternatives do not include costs for backup capacity.

- No allowance is made for the possible need for grease/oil removal prior to the BAF units. Indications are that yearly cleaning of the cells will suffice.
- Primary effluent pumping facilities are assumed with a 15 foot static lift for the activated sludge and RBC alternatives and a 20 foot static lift for the TF/SC alternative.
- No costs are included for administration building or laboratory facilities.
- In general, the cost estimates assume open tankage without covers. For the BAF alternative the filter surface area will be open with pipe galleries and service areas enclosed. For the RBC process insulated fibreglass covers are included for the RBC units.
- Land costs are not included in the estimates. Land area requirements are generally the same for all alternatives with the exception of the BAF process. The land area south of Ojibway Parkway will not be required for treatment facilities if the BAF process is selected.
- The cost estimates do not include any allowance for engineering, contingencies, taxes or interim financing.
- Capital cost estimates are in terms of 1992 dollars. The Engineering News Record Construction Cost Index has been used to relate estimates from other areas and previous years to a mid-1992 Toronto CCI value of 6730.

The accuracy of the cost estimates presented in Appendix J has also been assessed. The level of confidence in capital cost estimates increases as a project progresses from a conceptual stage to a detailed design state. Typically, the degree of accuracy at an early stage may be in the range of ±30 percent or greater becoming more refined at a level of ±10 percent at the end of the final design phase. Given the degree of detail possible in this study, it is estimated the level of accuracy of the capital cost presented are in the range of ±20 percent.

The cost estimates are useful for comparison purposes and generally agree with process comparisons from literature sources. **Use of the estimates to predict actual construction costs for any of the processes is not recommended as a more detailed level of design would be required to develop costs taking into account site specific variables.**

The capital cost estimates developed for each of the processes tested indicates that each process would be less expensive than an equivalent conventional ASP. The cost comparison of each process with respect to an equivalent conventional ASP are shown in Table 4.5, in terms of percentage.

Table 4.5
Comparison of Capital Costs for Each Process Tested at the West Windsor Treatment Plant

Process	Percentage of Capital Cost to Conventional ASP Capital Cost (%)[1]	
	@ 159,000 m^3/d Capacity	@ 273,000 m^3/d Capacity
1. Conventional ASP	100	100
2. TF/SC process	87	87
3. RBC process - Conventional SORs on final clarifier	87	89
- SORs determined from pilot testing	72	75
4. BAF process[2]	83	89
5. Modified ASP (as tested at West Windsor)	76	76

Notes:
[1] Differences between percentage at 159,000 and 273,00 m^3/d capacity due to round off in tank sizing.
[2] Difference between percentage at 159,000 and 273,000 m^3/d due to additional primary clarifier added at 273,000 m^3/d plant capacity to handle backwash solids

Capital cost estimates for each process varied from 76 percent to 89 percent of an equivalent conventional ASP. Capital costs for the RBC process varied from 73 percent to 88 percent (average of 159,000 and 273,000 m^3/d capacity) depending on the clarifier loading (i.e SOR) used. Pilot testing indicated that the RBC process could meet the effluent objectives established for the West Windsor PCP at a SOR which is about twice that used for a conventional suspended growth ASP. Indications are that a full-scale RBC facility would have to demonstrate that it could perform at these higher SORs and therefore the City intends to carry the higher capital cost during this evaluation (i.e. 88 percent of a conventional ASP). Capital costs for the BAF process did not include the cost of additional cells required to account for cells being out-of-service for cleaning. During storm events additional cells will be required to maintain sufficient cells in service.

Operation and maintenance (O&M) costs were also estimated for comparison purposes. Of the O&M costs, only electrical energy costs have been used to develop a comparison of operational costs for the alternative treatment processes. The other costs components of O&M costs have been excluded for the following reasons:

- manpower costs are difficult to predict with any accuracy and are probably similar for each process
- natural gas and municipal water usage are expected to be similar
- polymer and iron or aluminum costs for phosphorus removal will be similar
- maintenance material costs are expected to be similar

For a further discussion of the estimated O&M costs see LaFontaine, Cowie, Buratto and Associates Ltd.'s report contained in Appendix J.

Electrical power usage was estimated to be the least for the TF/SC plant, since this plant has minimal aeration requirements. Electrical power usage for the RBC and BAF plants was similar. Electrical power usage for the ASP, either conventional or modified, was estimated to be the highest due to the aeration requirements. On a net present value basis, the four processes tested are less costly than an equivalent ASP. The four processes vary from 79 percent to 89 percent of a conventional ASP on a net present value basis.

Overall, each of the four processes tested at West Windsor are expected to be more economical than a conventional activated sludge plant. However, the capital and O&M costs for each process are expected to be the same at this level of accuracy.

Section 5

OVERALL COMPARISON OF PROCESS ALTERNATIVES

To assist in determining the alternative treatment processes applicable to the West Windsor PCP, a comparison approach was used. This comparative approach emphasized the different attributes of each process. The results can be updated in the future to include additional information or different circumstances that may apply at the plant.

The approach compares the various processes using a number of main categories. Each category includes specific items. Each item was scored based on a similar scale to indicate the relative performance of each process. Items were weighted depending on the importance to West Windsor. Each item's rating was estimated based on objective operational performance (ie. effluent concentrations) or subjective opinion. To minimize the individual bias on the subjective opinion items and obtain a rounded opinion on each item, an evaluation sheet was sent to Steering Committee members to obtain their input. The comparison procedure is outlined in a Technical Memorandum sent to the project Steering Committee (Appendix I).

Each process was compared based on the following categories:

- Effluent performance
- Hydraulic performance
- Solids handling and solids yield
- Operational items
- Economics

The first three categories were compared based on the performance of all the pilot plants over the eleven-month period and included the performance of the BAF during the extension period. The fourth category, operational concerns, was assessed with the assistance of specific project Steering Committee members. Specific members individually scored the various items in selected categories on the evaluation sheet.

The economics category has been addressed by Lafontaine, Cowie, Buratto and Associates (LCBA) and details are summarized in Appendix J. The results for this category are based on the estimates for the capital costs of each process based on the pilot plant's performance. At the present time, LCBA indicates that each process would have a similar capital cost, within +/- 20 percent. For this report, the capital cost of each process, as well as the operating and maintenance and the net present worth, were assumed to be the same for each process.

Tables 5.1 and 5.2 show the main categories, the items in each category, and the weighting factor that was assigned to each item. The weighting factor emphasizes the items most important to the upgrade at the West Windsor PCP. Thirty-eight items were used to compare the processes. Of the main categories, 36 percent of the

Table 5.1
Results of Rating Comparison of Pilot Plant Processes Using 2 mg/L Effluent Ammonia-N Target

Process	Category 'A' Effluent Performance														Cat 'B' Hyd. Perf.		Cat 'C' Solids Prod			Category 'D' Operational Concerns																Category 'E' Economics			Totals					
	MONTHLY COMPLIANCE					PROB. LESS THAN OBJECTIVE					PROB LESS THAN UPPER																																	
	TSS	TBODS	TP	NH3-N	UN-IONIZED NH3-N	TSS	TBODS	TP	NH3-N	UN-IONIZED NH3-N	TSS	TBODS	TP	EFF. TOXICITY	DIURNIAL Q	PEAK DAY Q	SOLIDS YIELD	SETTLING CHARACTERISTICS	DEWATER CHARACTERISTICS	OPERATIONAL BASE	OPERATIONAL FLEXIBLITY	ENVIRONMENTAL CONCERNS	KNOWLEDGE OF PROCESS	PLANT SIZE	EFFLUENT AESTHETICS	CONSTRUCTION IN STAGES	STARTUP BASE	CONTROL	AUTOMATION	SEASONAL LOADINGS	SAMPLING REQUIREMENTS	OPERATIONAL PROBLEMS	MAINTENANCE PROBLEMS	NOISE AND ODOUR	POTENTIAL FOR CONVERSION TO BNR	CAPITAL COST	O&M COST	NET PRESENT WORTH	CATERGORY A	CATERGORY B	CATERGORY C	CATERGORY D	CATERGORY B	GRAND TOTAL
WEIGHTING	5.5	5.5	5.5	3.0	5.5	1	1	1	1	1	1	1	1	3	3	4	3	3	3	2	2	1	1	1	1	1	1	1	1	1	1	1	1	1	1	10	10	10						
100 %	55	55	55	30	55	10	10	10	10	10	10	10	10	30	30	40	30	30	30	20	20	10	10	10	10	10	10	10	10	10	10	10	10	10	10	100	100	100	360	70	90	180	300	1000
TF/SC	7	5	8	10	10	5	5	6	8	10	9	9	10	10	10	10	10	10	10	6	9	9	7	5	7	9	8	8	9	8	7	7	8	8	2	10	10	10						
	39	28	44	30	55	5	5	6	8	10	9	9	10	30	30	40	30	30	30	12	18	9	7	5	7	9	8	8	9	8	7	7	8	8	2	100	100	100	287	70	90	132	300	879
RBC	10	10	10	4	10	10	8	9	5	10	10	10	10	10	10	10	10	10	10	10	7	9	7	7	5	9	9	7	9	9	9	9	8	8	2	10	10	10						
	55	55	55	12	55	10	8	9	5	10	10	10	10	30	30	40	30	30	30	20	14	9	7	7	5	9	9	7	9	9	9	9	8	8	2	100	100	100	334	70	90	141	300	935
BAF	10	10	10	4	10	10	9	10	5	10	10	10	10	10	10	10	10	10	10	7	10	8	6	10	10	8	9	10	10	10	7	7	6	8	2	10	10	10						
	55	55	55	12	55	10	9	10	5	10	10	10	10	30	30	40	30	30	30	14	20	8	6	10	10	8	9	10	10	10	7	7	6	8	2	100	100	100	336	70	90	145	300	941
ASP	8	8	8	10	10	7	7	8	9	10	9	9	10	10	10	9	10	10	10	7	9	9	10	6	9	8	7	8	9	7	7	7	8	8	7	10	10	10						
	44	44	44	30	55	7	7	8	9	10	9	9	10	30	30	36	30	30	30	14	18	9	10	6	9	8	7	8	9	7	7	7	8	8	7	100	100	100	316	66	90	142	300	914

Table 5.2
Results of Rating Comparison of Pilot Plant Processes Using 3 mg/L Effluent Ammonia-N Target

Process	Category 'A' Effluent Performance														Cat 'B' Hyd. Perf.		Cat 'C' Solids Prod			Category 'D' Operational Concerns																Category 'E' Economics			Totals					
	MONTHLY COMPLIANCE					PROB. LESS THAN OBJECTIVE					PROB LESS THAN UPPER																																	
	TSS	TBOD5	TP	NH3-N	UN-IONIZED NH3-N	TSS	TBOD5	TP	NH3-N	UN-IONIZED NH3-N	TSS	TBOD5	TP	EFF. TOXICITY	DIURNAL Q	PEAK DAY Q	SOLIDS YIELD	SETTLING CHARACTERISTICS	DEWATER CHARACTERISTICS	OPERATIONAL EASE	OPERATIONAL FLEXIBILITY	ENVIRONMENTAL CONCERNS	KNOWLEDGE OF PROCESS	PLANT SIZE	EFFLUENT AESTHETICS	CONSTRUCTION IN STAGES	STARTUP EASE	CONTROL	AUTOMATION	SEASONAL LOADINGS	SAMPLING REQUIREMENTS	OPERATIONAL PROBLEMS	MAINTENANCE PROBLEMS	NOISE AND ODOUR	POTENTIAL FOR CONVERSION TO BNR	CAPITAL COST	O&M COST	NET PRESENT WORTH	CATERGORY A	CATERGORY B	CATERGORY C	CATERGORY D	CATERGORY E	GRAND TOTAL
WEIGHTING	5.5	5.5	5.5	3.0	5.5	1	1	1	1	1	1	1	1	3	3	4	3	3	3	2	2	1	1	1	1	1	1	1	1	1	1	1	1	1	1	10	10	10						
100 %	55	55	55	30	55	10	10	10	10	10	10	10	10	30	30	40	30	30	30	20	20	10	10	10	10	10	10	10	10	10	10	10	10	10	10	100	100	100	360	70	90	180	300	1000
TF/SC	7	5	8	10	10	5	5	6	9	10	9	9	10	10	10	10	10	10	10	6	9	9	7	5	7	9	8	8	9	8	7	7	8	8	2	10	10	10						
	39	28	44	30	55	5	5	6	9	10	9	9	10	30	30	40	30	30	30	12	18	9	7	5	7	9	8	8	9	8	7	7	8	8	2	100	100	100	288	70	90	132	300	[illegible]0
RBC	10	10	10	9	10	10	8	9	7	10	10	10	10	10	10	10	10	10	10	10	7	9	7	7	5	9	9	7	9	9	9	9	8	8	2	10	10	10						
	55	55	55	27	55	10	8	9	7	10	10	10	10	30	30	40	30	30	30	20	14	9	7	7	5	9	9	7	9	9	9	9	8	8	2	100	100	100	351	70	90	141	300	952
BAF	10	10	10	8	10	10	9	10	7	10	10	10	10	10	10	10	10	10	10	7	10	8	6	10	10	8	9	10	10	10	7	7	6	8	2	10	10	10						
	55	55	55	24	55	10	9	10	7	10	10	10	10	30	30	40	30	30	30	14	20	8	6	10	10	8	9	10	10	10	7	7	6	8	2	100	100	100	350	70	90	145	300	955
ASP	8	8	8	10	10	7	7	8	10	10	9	9	10	10	10	9	10	10	10	7	9	9	10	8	9	8	7	8	9	7	7	7	8	8	7	10	10	10						
	44	44	44	30	55	7	7	8	10	10	9	9	10	30	30	36	30	30	30	14	18	9	10	6	9	8	7	8	9	7	7	7	8	8	7	100	100	100	317	66	90	142	300	915

weighting was attributed to the effluent performance of the processes. Weights of seven and nine percent went toward the process hydraulic performance and solids handling aspects, respectively. Eighteen percent of the evaluation was assigned to the operational items associated with each process. Thirty percent was assigned to the estimated cost of each process for the West Windsor site.

The comparative results for the performance of each process under the first three categories were discussed at the end of Section 3. In Section 3, the performance in terms of meeting the monthly design objective, upper limit value, effluent toxicity, hydraulic performance, and solids handling was summarized. From this discussion, each process was compared based on a score between 0 and 10. Appendix I summarizes details on the method used to score these categories.

The Steering Committee members' input for Category D is summarized in Table 5.1 and Table 5.2. Eight completed questionnaires were returned to CH2M HILL for use in scoring this category. The overall scoring, for each item in this category was based on the arithmetic average of the questionnaire results.

Table 5.1 shows the completed comparison table with the scoring for each process based on a target effluent ammonia-N concentration of 2 mg/L. Table 5.2 shows the same comparison table, based on a target effluent NH_3-N concentration of 3 mg/L. The differences in the performance of each process were evident for Categories A, B, and D. For categories C and E, each process was rated the same. Category C was considered the same for each process based on the results of the solids handling bench-top testing (Section 3). Category E has been assumed the same for this comparison based on the cost estimated by LCBA.

For Category A, effluent performance, the RBC and BAF processes scored the highest based on overall consistent effluent quality, including nitrification. The RBC's performance during the winter with respect to nitrification was reduced due in part to the intentional operation of the pilot plant. In the winter, when the NH_3-N concentration increased above 2 mg/L, consideration was given to reducing the loading. However, because the unit was meeting the un-ionized NH_3-N objective, changes in the loading were not made. Table 5.2 indicates the process nitrified well with respect to meeting an effluent ammonia-N target of 3 mg/L. If the loading had been reduced, further nitrification would be expected.

The BAF process initially ranked lower than the RBC and ASP due to the process interruption associated with nitrification. However, during the extension period the BAF was able to nitrify through the winder period. Therefore, this process rates the same or better than the RBC unit. The BAF unit was rated slightly higher than the RBC because of its enhanced effluent quality in terms of SS, BOD_5 and TP.

Numerically, the ASP process was scored second due to consistent nitrification from the pilot plant. The ASP did not perform as well as the RBC and the BAF units with respect to the effluent SS, total BOD_5 and TP objectives. The performance of the ASP unit may be partly due to the size of the unit tested.

The TF/SC was scored the lowest due to the process's apparent inability to meet design objectives on a monthly basis, especially for SS, total BOD_5 and TP.

The operation of the SC was further evaluated during the second year of operation with no improvement in effluent performance. To compare the excellent particulate and associated contaminant removal obtained by the BAF, the spread of the effluent SS concentrations from each process was determined (Table 5.3). The spread of each process' effluent SS concentration was compared by investigating the percentage of time on a daily basis that the effluent SS concentration from each process was less than or equal to 5, 10, 15, and 25 mg/L. Table 5.3 is based on the period of time when each process was operating as designed or after optimizing each process, during the first year of operation when all pilots were operated and also includes the data available for the extension period (TF/SC and BAF only). The results show that the BAF effluent SS concentration was less than 5 mg/L for 71 percent of the time. The effluent was less than 10 and 15 mg/L for 95 and 98 percent of the time, respectively. This performance is significantly better than any of the other units' performance. The RBC pilot plant produced the next best results with less than the assigned concentrations for 28, 85, and 96 percent of the time. The ASP and the TF/SC pilot plants' effluent SS concentrations exceeded 10 mg/L the majority of the time.

Table 5.3
Performance of Pilot Plants With Respect to Effluent SS on a Daily Basis

Process	**SS Concentration Probability Less Than or Equal To**			
	5 mg/L	**10 mg/L**	**15 mg/L**	**25 mg/L**
TF/SC	4	23	55	92
RBC	28	85	96	99
BAF	71	95	98	99
ASP	17	41	75	92

For Category B, hydraulic performance, the RBC and TF/SC were scored the same, since neither of these processes had any difficulty handling peak day and diurnal flow rates. The ASP process was scored slightly lower, since this process had difficulties at pilot scale handling high hydraulic loading. This process is expected to have no difficulty handling peak day and daily diurnal flows at full-scale. The Little River PCP (ASP) was designed to treat peak day and diurnal flows, and has been shown to handle both. Therefore, the rating for this process was only dropped slightly. During the first year of operations there were concerns with the BAF handling peak flows. The hydraulic capacity of this unit was addressed during the extension and the unit was able to handle a 2.2 peak factor for a 3- to 4-day period. Hydraulic limitations in the BAF appear to have been addressed by periodic air scours.

Category D summarizes the results of the evaluation sheet filled out by the Steering Committee members. Not all evaluation sheets were filled out for each item. Committee members (if they filled out the evaluation) rated the items which they felt confident rating. Based on the average results, the RBC, BAF, and ASP were rated the highest. The performance of the BAF process in terms of operational concerns, was increased due to further evaluation of this process, during the extension period. In particular, the BAF's ability to handle seasonal loading and provide flexible operations was deemed to be better than the other processes, since this process has the ability to put cells into service and immediately obtain mechanical filtration. The TF/SC process was rated the lowest; however, it was rated within 9 percent of the best process.

All of the process units had similar overall scores based on the analysis method and scores assessed. Despite this similarity, there may be decisive factors (eg. effluent achievable) that override small differences in this comparative evaluation. Thus, the comparative evaluation should only be used as an additional tool to assess plant applicability to the City of Windsor.

Based on the overall comparison, the BAF process scored highest for the upgrade of the West Windsor PCP. The RBC and ASP were within 0.5 and 3.6 percent of the BAF's performance, respectively, based on an average of the two effluent ammonia-N criteria. Each of the three highest rated processes has advantages and disadvantages that will need to be evaluated by the City before a selection of a process can be made. On an overall basis, the main advantages for each of the recommended process are:

RBC process:	results of pilot-scale indicate this is an easy process to operate and may be designed with high final clarifier SORs.
BAF process:	provides enhanced particulate removal and should more stringent effluent requirement, such as tertiary treatment, be mandated at West Windsor, this process would have a clear advantage
ASP:	provides a process that is well known to the City and a process that operates efficiently at the Little River PCP

The TF/SC process would be a possible process for the City; however, based on its performance with the current high quality primary effluent combined with the relatively dilute raw sewage, it rates below the other processes tested. This conclusion is especially relevant in terms of the TF/SC's effluent performance on a consistent monthly basis.

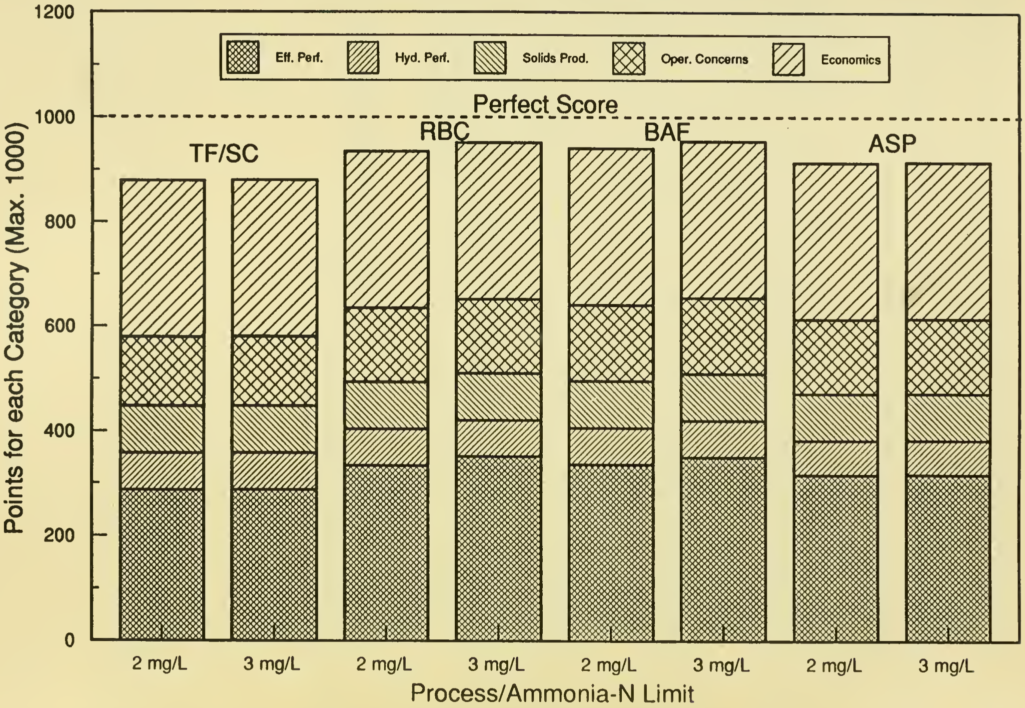
Figure 5-1 shows a bar chart of the scoring results of each process at both ammonia-N targets. Although all the processes rated within 8 percent of each other at either target NH_3-N concentration, the TF/SC monthly effluent performance, especially for SS, total BOD_5, and TP was less consistent than the other processes. Based on the monthly compliance for SS, total BOD_5 and TP, the RBC and BAF pilot plants met the monthly objective every month while the TF/SC only met these three objectives about two-thirds of the time.

The coupled TF/SC process appears to require a different operational mode to generate sufficient solids inventory in the SC chamber. Despite values near the design objective, limitations in control of effluent particulates, even at reduced clarifier surface overflow rates, appears to limit the biomass inventory that can be maintained in the SC chamber. This limitation became apparent only after considerable operating experience was achieved and solids inventory calculations (estimated mass balance) were practical. With the dilute raw sewage and enhanced primary effluent quality achieved in the full-scale plant, the SC process will be susceptible to biomass inventory and plant upset problems. Future work on the TF/SC should focus on generating an increased solids load and accumulation or biomass growth rate in the SC. In particular, it is recommended that the organic concentration to the TF unit be investigated to achieve higher biomass loadings (sloughed TF solids) to the SC. Alternatively, the ability to operate a single-step carbon removal-nitrification tower with secondary clarification with or without the assistance of a polymer should be assessed.

The TF/SC process required the full-scale plant to reduce its treatment, in terms of reducing or eliminating polymer dosage, to operate as designed. Therefore, if the City chooses to make the greatest use of the existing plant's enhanced primary treatment capability, the TF/SC should not be considered as a possible alternative for the plant upgrade.

Any of the other three processes, and especially the RBC and BAF, take advantage of the current enhanced primary treatment to produce an effluent quality that met the effluent objectives of this study.

Figure 5-1 - Comparison of Process Alternatives for the Upgrade of the West Windsor PCP



Section 6
CONCLUSIONS AND RECOMMENDATIONS

The existing West Windsor PCP will require upgrading to the equivalent of secondary wastewater treatment standards when these standards are promulgated by the MOE MISA program. Effluent design objectives for an expanded plant are anticipated to be:

- Total BOD_5 and SS = 15 mg/L
- TP = 0.5 mg/L
- Total un-ionized NH_3-N = 0.1 mg/L

Target effluent ammonia-N concentrations of 2 and 3 mg/L were also used.

To determine potential upgrading alternatives, three pilot scale and one bench scale process were operated at the West Windsor PCP, including:

- the combined trickling filter/solids contact (TF/SC) process
- the rotating biological contactor (RBC) process
- the biological aerated filter (BAF) process
- the modified activated sludge process (ASP; at bench scale)

The pilot plant testing was conducted over an eleven-month period starting in July 1990. Two processes, the TF/SC and BAF, were operated for an additional year long period to investigate additional concerns

The following conclusions and recommendations are made based on the performance of the four pilot plants.

CONCLUSIONS

An evaluation of the performance of each of the pilot plants is presented in Section 5. The overall performance of each process as well as other pertinent conclusions are summarized below.

GENERAL

- All four pilot plants used the existing effluent from the full-scale West Windsor PCP clarifiers No. 5 and 6. The pilot plant influent concentration averaged 22, 41, 0.8 and 8.9 mg/L for SS, total BOD_5, TP and total NH_3-N for the eleven-month period. The influent concentrations were increased for the period October 30, 1990 to January 11, 1991 by discontinuing the polymer addition to primary clarifiers No. 5 and 6. During this period without polymer, the influent SS and TP to the pilot plants increased to 31 and 1.2 mg/L. The total BOD_5 concentration increased to 46 mg/L.

- The TF/SC and BAF processes were operated for an additional year long period. During the initial extensions period (June 1, 1991 to August 2, 1991) the pilot plant influent concentration averaged 25, 43, 1.6, and 9.6 mg/L for SS, total BOD_5, TP, and total NH_3-N while using ferric only. During the follow-up extension (September 17,1991 to April 26, 1992) the pilot plant influent concentration averaged 24, 47 0.6, and 9.9 mg/L, while using ferric and polymer.

TF/SC

- The coupled TF/SC pilot plant was continuously operated from December 16, 1990 to May 1, 1991, from June 1 to August 2, 1991, and from September 17, 1991 to April 26, 1992. The TF operated on its own in conjunction with a final clarifier from August 30 to November 4, 1990. Attempts were made to increase the TF loading prior to and during that time in an attempt to generate a sustainable biomass in the SC unit.

- During the period when the TF was operated on its own, the process's performance was the same as the coupled TF/SC's performance in terms of effluent SS and TP. The TF/SC process performed better than the TF alone in terms of effluent total BOD_5 and NH_3-N. The SOR of the final clarifier during the period when the TF was operated on its own was 55 $m^3/m^2 \cdot d$, which is greater than three times the average SOR that was used when the combined TF/SC was in operation. A parallel comparison of the TF/SC and TF process during the follow-up extension period confirmed these results.

- During the period when the TF/SC was operating, the unit met the monthly design objectives for 66, 50, 83 and 100 percent of the time for SS, total BOD_5, TP, and un-ionized NH_3-N. The TF/SC effluent total NH_3-N was less than 2 mg/L for 100 percent of the time on a monthly basis.

- During the extension period attempts were made to improve the performance of the SC by increasing the loading to the TF tower, reducing the SC volume and operating the SC with a solids reaeration pass. None of these actions improved the performance of the SC portion of the TF/SC process.

RBC

- The RBC pilot plant operated continuously from the beginning of July 1990 to the end of May 1991.
- During this period the process met the monthly design objectives for SS, total BOD_5, TP and un-ionized NH_3-N for 100 percent of the time. The effluent total NH_3-N was less than 2 and 3 mg/L for 40 and 90 percent of the time on a monthly basis.
- Total NH_3-N concentrations greater than 2 mg/L resulted due to nitrogen loadings which were intentionally maintained at loadings exceeding complete nitrification capability as this was not necessary to meet the target design objectives.

BAF

- The BAF pilot plant operated from August 20, 1990 to August 2, 1991, and from September 17, 1991 to April 26, 1992.
- During this period the process met the monthly design objectives for SS, total BOD_5, TP and un-ionized NH_3-N for 100 percent of the time. The process attained an effluent total NH_3-N concentration of 2 mg/L for 25 percent of the time for the first period of operations. During the follow-up extension period after the media depth was increased to 2.5 m, the BAF process nitrified to an average effluent ammonia-N concentration of 3 mg/L. Overall, the effluent total NH_3-N was less than 2 and 3 mg/L for 40 and 80 percent of the time on a monthly basis.
- The BAF process generated a highly polished effluent.
- Nitrification was interrupted in the late Fall of 1990 due to a number of vigorous backwashes which were conducted to investigate potential hydraulic bottlenecks in the process. Once nitrification was lost, attempts were made to re-seed the unit, but these efforts were unsuccessful at cold wastewater temperatures. The unit began nitrifying again in the Spring of 1991. Prior to the interruption of nitrification, the BAF effluent total NH_3-N averaged 1.9 mg/L. During the extension period nitrification was re-established, especially after the media depth was increased to 2.5 m.
- Hydraulic problems associated with handling peak day and diurnal flows were investigated and corrected by making operational changes. During the follow-up extension period the BAF was able to handle a flow of 5.5 $m^3/m^2 \bullet h$.

ASP

- The ASP bench-top pilot plant was operated from November 16, 1990 to May 31, 1991. The unit was out-of-service for a two-week period during the winter due to cold weather operational mechanical problems.

- The ASP unit obtained effective nitrification, meeting the monthly effluent un-ionized NH_3-N objective for 100 percent of the time. The ASP met the other effluent objectives for SS, total BOD_5 and TP for over 83 percent of the time on a monthly basis. The effluent total NH_3-N concentration was less than 2 mg/L for 100 percent of the time on a monthly basis.

SOLIDS PRODUCTION AND HANDLING

- The solids yields from each of the alternative processes were similar and within the range expected for a secondary biological treatment process (0.6 to 0.8 kg VSS/kg BOD_5 removed).

- The solids handling aspect of each process was evaluated by conducting a series of bench-top settling and dewaterability tests. The results indicated that the waste solids from these processes should not present particular difficulties in connection with thickening and dewatering alone or combined with primary sludge. The ASP unit did not generate sufficient waste solids to be tested; however, the City of Windsor's Little River PCP handles waste activity sludge without difficulty.

CONCLUSIONS SUMMARY

- A comparison of advantages and disadvantages is provided in Table 6.1.

- The recommended process loadings based on the performance of the pilot plants are shown in Table 6.2. In general, the loading rates are close to the initial target rates used at startup (Appendix B). Loading rates for the SC unit of the TF/SC process are not available because during the testing the optimal loading for the unit could not be determined. The SC loadings appeared to be near the limit of adequate operation due to difficulties maintaining solids inventory and the solids loss in May 1991. The nitrogen loading rates for the RBC and BAF units have been calculated to meet an effluent total NH_3-N concentration of 3 mg/L. The requirement to achieve this effluent total NH_3-N, particularly in cold weather operation, should be verified with the MOE as it was not a design objective in this project and has a significant impact on plant sizing requirements.

Table 6.1
Overall Summary of Pilot Plant Processes

Process	Advantages	Disadvantages
TF/SC	• Nitrification consistent through TF tower • TF easy to operate	• Marginal effluent with respect to SS, total BOD_5 and TP concentrations compared to objective • Difficulty maintaining solids inventory in SC tank
RBC	• Consistent effluent with respect to SS, total BOD_5 and TP compared to objective • Nitrification maintained • Easy to operate and maintain • good settling sludge; pilot operated at an elevated SOR • simple process	• Aesthetics of effluent is poor
BAF	• Excellent effluent with respect to SS, total BOD_5 and TP; tertiary quality • Good visual aesthetics • Nitrification possible with 2.5 m media depth	• More complex mechanical/electrical process • Nitrification affected by excessive backwash and peak day flow events
ASP	• Consistent nitrification • System well known to City	• Effluent quality was not as good with respect to effluent SS, total BOD_5, and TP compared to RBC and BAF processes. Effluent did meet these limits 83 percent of the time on a monthly basis. Some of the units performance was due to the size of the unit.

Table 6.2
Recommended Process Loading Rates Based on Pilot Plant Studies

Process			
	Organic	Nitrogen	Hydraulic
TF/SC			
A) TF Tower	0.3 kg $BOD_5/m^3 \cdot d$	0.08 kg NH_3-$N/m^3 \cdot d$	36 $m^3/m^2 \cdot d$
B) SC	NA	NA	NA
C) Clarifier (peak)	--	--	29.4 $m^3/m^2 \cdot d$ (peak)
RBC			
A) Disks	3.5 g $BOD_5/m^2 \cdot d$	1.2 g $TKN/m^2 \cdot d^*$	71.5 $m^3/m^2 \cdot d$ (peak)
B) Clarifier (peak)	--	--	
BAF	1.2 kg $BOD_5/m^3 \cdot d$	0.22 to 0.30 kg $TKN/m^3 \cdot d^{**}$	132 $m^3/m^2 \cdot d$ (peak)
ASP			
A) Aeration	0.32 kg $BOD_5/m^3 \cdot d$	0.06 kg NH_3-$N/m^3 \cdot d$	3.9 h
B) Clarifier (peak)			29.4 $m^3/m^2 \cdot d$

Notes:
*Based on using the linear regression of NH_3-N loading and effluent NH_3-N concentrations to obtain an effluent of 2 mg/L.
**To be verified for cold temperature operation.

- The economic evaluation is based on comparative full-scale process cost estimates developed by LaFontaine, Cowie, Buratto and Associates Ltd. Overall, each of the four processes tested at West Windsor are expected to be more economical than a conventional activated sludge plant. Capital costs for each process varied from 76 to 89 percent of an equivalent conventional ASP. However, the capital and O&M costs for each process are expected to be the same at this level of accuracy.

COMPARATIVE EVALUATION

- Overall based on an evaluation including effluent performance, hydraulic performance, solids handling, operational concerns and economics, the BAF process scored highest as an alternative for the West Windsor PCP upgrade. The evaluation results for the RBC and ASP units were marginally less than the BAF. The BAF has the advantage of obtaining an effluent of enhanced tertiary quality. The TF/SC rated the lowest, although not significantly below the other three processes based on a summary of all categories.

- Although the combined TF/SC did not rate significantly below the other three processes based on a summary of all categories, it did not perform as well as the other processes in terms of monthly effluent SS, total BOD_5 and TP. Table 6.3 summarizes the results of the comparative evaluation of each process in terms of effluent performance. The rating is based on a possible total score of 360 and evaluates effluent performance on a monthly and daily basis. When based as a percentage (i.e. 360 being 100 percent) the TF/SC performed below the average of all four processes regardless of whether an effluent ammonia-N target of 2 or 3 mg/L is used. In addition, the reliability or need for a solids contactor with the TF tower is in question due to the dilute raw sewage and effective primary treatment at the West Windsor PCP. Given the difficulty in maintaining the SC process and the solids washout which occurred in May 1991, the coupled TF/SC process is not recommended for the City of Windsor without additional testing with different influent loading conditions or process configurations.

Table 6.3
Summary of Pilot Plants' Effluent Performance

Process	Effluent Performance			
	Rating Points (Out of a Possible 360)		Percent of Maximun (%)	
	NH_3-N Target		NH_3-N Target	
	2 mg/L	3 mg/L	2 mg/L	3 mg/L
TF/SC	287	288	80	80
RBC	334	351	93	98
BAF	336	350	93	97
ASP	316	317	88	88

RECOMMENDATIONS

As a result of conducting the pilot studies at the West Windsor PCP, the following recommendations are made:

- The coupled TF/SC process, based on its performance during the pilot testing and with dilute influent concentrations, is not recommended to upgrade the West Windsor PCP.
- If the coupled TF/SC or TF evaluation is continued , additional testing should include an increased organic loading to the TF unit. This would require a trade-off between decreased (or no) chemical consumption in primary treatment followed by the need for additional TF/SC secondary treatment capacity.
- If the City decides to proceed with either the TF/SC (or other TF configurations) BAF or RBC processes further site inspections and plant data should be reviewed to assist in familiarization with these processes.
- That, to obtain a complete understanding of the performance and viability of the BAF process, the City inspect a number of full-scale BAF facilities which have been operating for several years. This is essential to gain a full understanding of:
  - capital and operating cost implications
  - manpower staffing needs
  - maintenance requirements
  - nitrification performance
  - operating strategies

The BAF process is an innovative process in North America and long-term operating experience is unavailable here. Operating facilities with a history of operation are located at:

- North-Bierley, Britain
- Lunéville, France
- Soissons, France

Section 7
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Section 8
GLOSSARY OF TERMS

Glossary of Terms

Term	Description
ABF	Activated Biofilter
Al^{+3}	Aluminum Ion
ASP	Activated Sludge Process
AVG	Average
BAF	Biological Aerated Filter
BATEA	Best Available Treatment Economically Achievable
BF/AS	Biofilter/Activated Sludge
Biocarbone	BAF supplier's process name
BOD_5	Biological oxygen demand
CALL	CANVIRO Analytical Laboratories Ltd.
CAT	Category
$CBOD_5$	Inhibited or Carbonaceous five-day BOD
CCI	Construction Cost Index
cfm	Cubic feet per minute
COD	Chemical oxygen demand
CONC	Concentration
CST	Capillary suction time
d	Depth
DAF	Dissolved air flotation
dia	Diameter
DIFF	Difference
DM	Dry material
DO	Dissolved oxygen
EFF	Effluent

Glossary of Terms

Term	Description
Fe^{+3}	Ferric Ion
F/M	Food-to-Microorganisms Ratio
HRT	Hydraulic retention time
INF	Influent
$kg/m^2 \cdot d$	Kilograms per square meter per day
$kg/m^3 \cdot d$	Kilograms per cubic meter per day
kW	Kilowatts
L	Length
LCBA	LaFontaine, Cowie, Buratto and Associates
LRPCP	Little River Pollution Control Plant
L/min	Litres per minute
m^3/d	Cubic meters per day
$m^3/m^2 \cdot d$	Cubic meter per square meter per day
$m^3/m^2 \cdot h$	Cubic meter per square meter per hour
MAX	Maximum
mg	Milligrams
mg/L	Milligram per litre
MIGD	Million imperial gallons per day
MIN	Minimum
MISA	Municipal-Industrial Strategy for Abatement
mL	Millilitres
mL/mo	Megalitres per month
MLSS	Mixed liquor suspended solids
mo	Month
MOE	Ministry of the Environment
Munters	TF media supplier

Glossary of Terms	
Term	**Description**
N	Nitrogen
NA	Not Available
NH_3-N	Ammonia nitrogen
NO_2-N	Nitrite Nitrogen
NO_3-N	Nitrate Nitrogen
OTV	Compagnie Generale des Eaux
O&M	Operations and maintenance
PCP	Pollution Control Plant
PF	Peak factor
PROB	Probability
PVC	Polyvinyl Chloride
QA/QC	Quality Assurance/Quality Control
RAS	Return activated sludge
RAW	Raw sludge or sewage
RBC	Rotating Biological Contactor
RF/AS	Roughing Filter/Activated Sludge
SC	Solids contactor
SLR	Solids loading rate
SOR	Surface overflow rate
SRT	Solids retention time
SS	Suspended solids
STD DEV	Standard deviation
SVI	Sludge volume index
SWD	Side wall depth
TAL	Total ammonia-N loading
$TBOD_5$	Total five-day biochemical oxygen demand

Glossary of Terms	
Term	**Description**
TF	Trickling filter
TF/AS	Trickling Filter/Activated Sludge
TF/SC	Trickling Filter/Solids Contactor
TKN	Total Kjeldahl nitrogen
TOL	Total organic loading
TP	Total phosphorus
t/mo	Tonnes per month
TS	Total solids
TSS	Total suspended solids
TVS	Total volatile solids
UV	Ultra-violet
V	Volume
VS	Volatile solids
VSS	Volatile suspended solids
W	Width
WAS	Waste activated sludge
WRP	Water Reclamation Plant
wrt	With respect to
WWPCP	West Windsor Pollution Control Plant
WWTP	Wastewater Treatment Plant

Appendix A

PROJECT REPORTS AND CHRONOLOGY OF PILOT PLANT OPERATIONS

Please contact Tony Ho (416-314-3877) for information/data contained in this Appendix.

Appendix B

INITIAL INVESTIGATION OF EACH TECHNOLOGY

CITY OF WINDSOR WEST WINDSOR PCP
UPGRADE

TECHNICAL MEMORANDUM NO. 1

TO: Steering Committee

FROM: CH2M HILL ENGINEERING LTD.

DATE: May 4, 1990

PROJECT: ONT29353.A0

SUBJECT: Review of Selected Treatment Alternatives for the West Windsor PCP

The following is a brief summary of the selected treatment alternatives to be pilot tested at the West Windsor PCP. The selected treatment alternatives are:

- Trickling Filter/Solids Contact Process
- Rotating Biological Contractor Process
- Biological Aerated Filter Process
- "Non-traditional" Activated Sludge Process

A brief summary of each process follows:

1.0 TRICKLING FILTER/SOLIDS CONTACT PROCESS

1.1 General

Trickling filters have historically been a popular treatment process in the U.S., due in part to their simplicity, reliability, stability, economy, ease of operation and energy-saving features. However, since 1970 activated sludge processes (suspended growth system) have become more prevalent due to more stringent effluent requirements. Effluent concentrations for BOD_5 and SS from trickling filter plants are often in the 20 to 40 mg/L range.

The discovery of the trickling filter/solids contact process (TF/SC) has resulted in an improved system which can achieve more stringent effluent requirements and/or operate at higher trickling filter loading rates, including partial or full nitrification at some sites. The typical TF/SC plant includes:

- a trickling filter,
- an aerobic solids contact period,
- a flocculation period, and

- a secondary filter.

Two critical operating features of the system are:

- solids must be maintained in an aerobic, flocculent state,
- solids must be recycled from the secondary clarifier to combine with trickling filter effluent.

The main function of the trickling filter is to reduce the soluble BOD_5 in the wastewater. The aerated solids contact chamber's primary purpose is to capture and flocculate solids and reduce particulate BOD_5.

The most common mode of TF/SC is the TF/SCR, or trickling filter/solids contact process with solids aeration. This system is depicted in Figure 1. This system along with the trickling filter, solids contact and flocculator clarifier uses a sludge reaeration tank (Figure 1 - MODE III). This particular configuration is suited to removal of a modest amount of soluble BOD_5 as well as particulate solids removal.

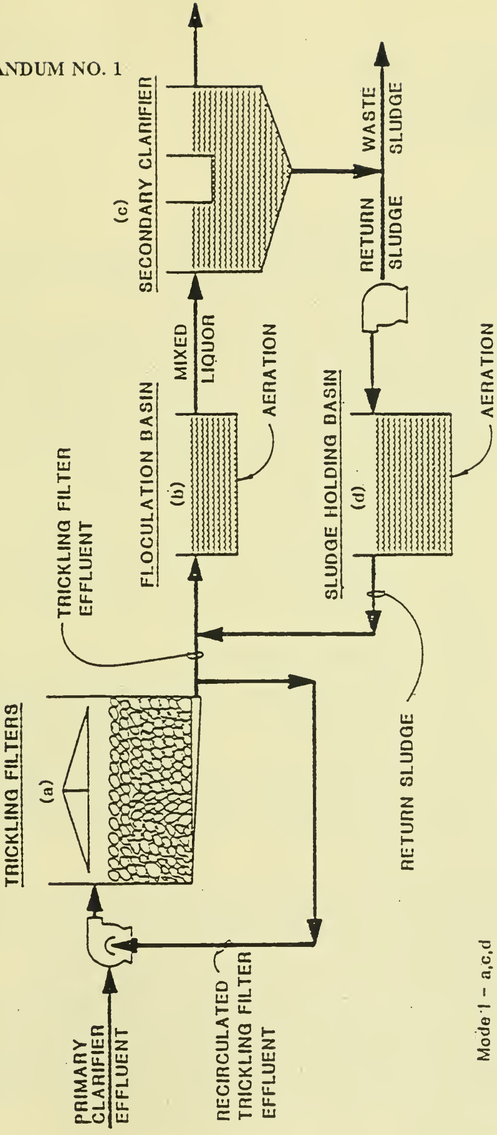
Two other configurations are used for more specific waste characteristics. For waste with a major fraction of soluble BOD_5, a TF/SC with no sludge reaeration is often used (Figure 1 - MODE II). For waste with a low soluble BOD_5 and high levels of particulate solids, a system with sludge reaeration and little or no solids contact might be used (Figure 1 - MODE I).

A review of a number of TF/SC plants performance for treated flows ranging from 1400 to 39,700 m^3/d resulted in effluent total BOD_5 and SS concentrations averaging 17 and 9 mg/L respectively. Carbonaceous BOD_5 concentrations from a few of these plants indicated the plants were partially nitrifying.

The trickling filter component of the TF/SC process is often designed based on the total organic loading (TOL) per volume of filter. Typcial loadings range from 0.64 to 0.96 kg $BOD_5/m^3 \cdot d$ utilizing plastic media. It is recommended that loadings should be reduced for trickling filters with rock media.

The aerated solids contact time is generally 1 hour or less based on the total flow, including the recycle flow rate. The solids retention time (SRT) in the aerated solids contact tank is generally less than 2 days. Fine or medium bubble diffusers are used to minimize turbulence in the activated solids contact tank while maintaining adequate dissolved oxygen levels.

The secondary clarifier is generally designed based on a clarifier overflow basis. Clarifier overflow rates of 16 $m^3/d \cdot m^2$ (400 (US) gpd/ft^2) for average dry weather flows are generally considered necessary to achieve an effluent quality of 10 mg/L total suspended solids in the effluent.



Mode I – a,c,d
Mode II – a,b,c
Mode III – a,b,c,d

Figure 1 – SCHEMATIC LAYOUT OF A TRICKLING FILTER/SOLIDS CONTACT PROCESS

1.2 Pilot Plant Requirements

To provide for a representative pilot test, a large pilot scale unit will be used. The average feed flow for this unit will be 175 m^3/d or 32 U.S. gpm. A schematic layout of the proposed pilot plant is shown in Figure 2.

In addition, to the TOL used to size the trickling filter, the hydraulic loading rate must also be considered. This is especially the case for a nitrification application. Munters, the media manufacturer, recommends a hydraulic loading of 29.4 $m^3/m^3 \cdot d$ (0.5 U.S. gpm/ft^2). In this case, the hydraulic loading or wetting rate is the criteria used to size the media volume. To maintain the full scale depth, to avoid poor pilot scale results, a pilot plant with 25.4 m^3 (896 ft^3) of media will be used. The dimensions are 2.4 m by 2.4 m surface area, with 4.3 m depth (8 x 8 x 14 ft).

For design of the pilot facility the following information form Munters was obtained:

- the media is self supporting, and will be laid in rows using standard 2 x 4 x 2 ft media sections. the rows alternate 90 degrees on each subsequent layer, to keep the flow directed into the media. At each row a thin blanket of liner (ie. PVC) will be required to divert flow back into the media.
- The media weight is 2 lb/ft^3 dry, and approximately 12 lb/ft^3 with biomass attached. West Windsor's biomass may be significantly lighter. since its loading is lower than the average application. However based on this weight, the expected weight of the loaded media is 4,880 kg (10,800 lb).
- Munters has indicated that ideally the trickling filter effluent should not be pumped, to avoid shearing the floc. However, due to size of the trickling filter, we have proposed pumping the trickling filter effluent, by a low shear type pump such as a progressive cavity or peristaltic type pump.

The remaining unit processes are discussed in Table 1. The loadings for each are within typical guidelines, except the clarifier overflow rate which exceeds that recommended. This limitation results due to equipment availability.

A description of the pilot facility depicted in Figure 2 follows. The feed flow will originate in a common feed well, from which a variable speed pump will lift the influent to the top of the trickling filter. The flow will be diverted into the trickling filter through a channel with a number of pipes attached, from which it will flow through a series of holes onto the filter media. The feed will be further broken up by attaching a splash plate under each hole. The media manufacturer had indicated that perfect wetting of the surface is not necessary since the media's cross flow channels will avoid short-circuiting through the system.

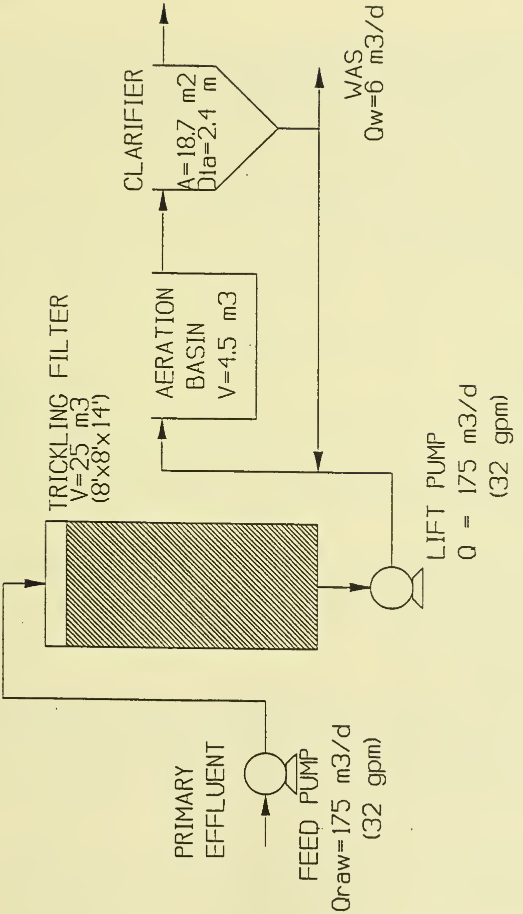


Figure 2 - SCHEMATIC LAYOUT OF TF/SC PILOT PLANT

Table 1
TF/SC PILOT PLANT LOADING CHARACTERISTICS

Parameter	Pilot Plant	Guideline
1 General		
• Flow (m^3/d)		
Avg.	175 (32 U.S. gpm)	
Peak	410 (75 U.S. gpm)	
• Influent Conc. (mg/L)		
$TBOD_5$	36	
TSS	26	
TKN	14.3	
2.Trickling Filter	V = 896 ft^3 = 25.4 m^3	
• TOL $\left(\frac{kg \cdot BOD_5}{m^3 \cdot d}\right)$	0.25	0.64 - 0.96 $\frac{kgBOD_5}{m^3 \cdot d}$
• Wetting Rate (hyd. loading $m^3/d \cdot m^2$)	29.4 (0.5 U.S. gpm/ft^2)	0.5 U.S. gpm/ft^2
3.Solids Contact Tank	V = 4.5 m^3	
• HRT (recycle included @ .25Q)	30 min	less than 60 min
• Recycle Rate	44 (8 U.S. gpm)	25% of Q_{RAW}
• SRT (Q_{WC} = 2.3 m^3/d)	2d	2d
4.Final Clarifier	37.5 $m^3/m^2 \cdot d$ (917 gpd/ft^2)	16 $m^3/m^2 \cdot d$ (400 gpd/ft^2)

The effluent from the trickling filter will be collected at the bottom of the filter and diverted into an approximately 300 gal tank with a maximum depth of 3 feet. A low shear pump will be used to lift the effluent approximately 10 feet into the aeration tanks. The aeration tanks will be two joined 900 gal tanks, each approximately half full. Both the trickling filter effluent and the return activated sludge flows will enter the first tank. Air will be supplied by fine bubble diffusers. The effluent will be taken from the second tank via a variable height stand pipe. Flow will proceed by gravity to the 8 foot diameter secondary clarifier. Effluent from the clarifier will flow along the ground to a manhole to be recycled back to the plant. Return flow will be pumped back to the aeration tank. Wasting will occur from the aeration tank either by a control valve by gravity or by batching through a pump. Wasting will also be diverted to the manhole. Waste will be temporarily collected in a tank (100 - 200 gal) to determine the volume and solid concentration.

2.0 ROTATING BIOLOGICAL CONTACTOR

2.1 General

The rotating biological contactor (RBC) is a fixed film wastewater treatment process which was developed and introduced on a commercial scale in Europe in the late 1950s and early 1960s. Application of RBCs in North America began only in the late 1960s and by 1975 there were still only a few small package plants operating in Canada. Today there are two full scale municipal facilities in Ontario (Guelph, Niagara Falls).

The RBC process involves a series of closely spaced discs anchored on a horizontal shaft that is supported above the surface of the wastewater in a semi-circular or rectangular tank. The shaft rotates exposing the biological growth on the disc surfaces alternately to the wastewater and atmosphere. The period the discs are exposed to the atmosphere provides the oxygen required for the anaerobic treatment of the wastewater.

A traditional approach used for design of RBCs is similar to that used for a suspended solids biological process, by defining the organic loading. A more common approach to design an RBC plant has been the use of substrate loading rate per unit area of disc. The organic load was found to be the variable that most closely correlated with the effluent BOD_5 concentration. The mass loading has also been proposed as the major parameter for designing nitrifying RBC systems.

Typical loadings stated by manufacturer's design manuals vary from 12 to 26 kg total $BOD_5/1000\ m^3 \cdot d$ for 90 percent removal of total BOD_5. For a system designed for combined BOD_5 removal plus nitrification (90 percent ammonia-N removal), the loading rate reduces to 8 to 17 kg total $BOD_5/1000\ m^2 \cdot d$. Actual plant operating data from six facilities varied from 4 to 18 kg total $BOD_5/1000\ m^2 \cdot d$, and resulted in substantial higher effluent total BOD_5 concentrations (14-40 mg/L).

Actual RBC operating data also indicated improved RBC performance at higher liquid temperatures. Pilot scale studies indicated a loading of 6.0 to 6.5 total $BOD_5/1000\ m^2 \cdot d$ would be selected to obtain an average effluent objective of 15 mg/L for BOD_5 and suspended solids at wastewater temperatures of 8 to 10 °C. To obtain nitrification (ie. TKN $\leq$ 5 mg/L), the pilot scale loading dropped significantly to 0.35 to 0.4 kg TKN/1000 $m^2 \cdot d$ for the same wastewater temperatures.

2.2 Pilot Plant Requirements

The RBC process is shown in Figure 3. The RBC unit has been obtained from Environment Canada. The RBC unit is a 2 m four stage unit, with a media surface area of 734 m^2. The RBC will be attached to a 1.6 m (5 ft) secondary clarifier also provided by Environment Canada.

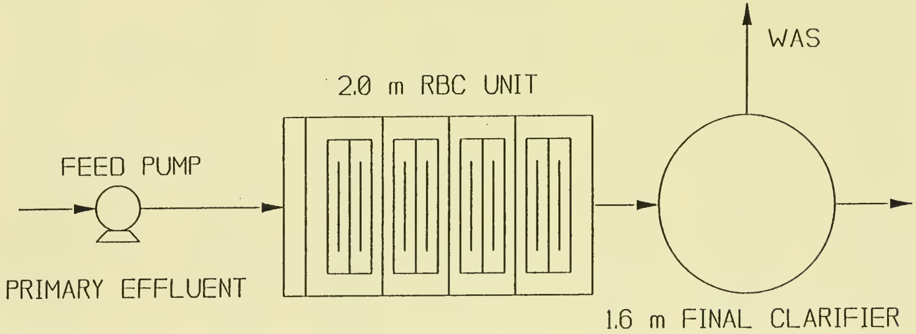
The pilot plant will be loaded at an average flow rate of approximately 65 m^3/d (12 U.S.gpm), with peak flows to 136 m^3/d (25 U.S. gpm). The resulting design guidelines for the RBC process are reviewed in Table 2. The results show that the loading criteria will be determined by the ammonia loading rate. The remaining parameters meet guideline or typical values for a RBC plant.

The pilot plant will obtain its feed flow from the common well located in the compost dryer building. The pump will have variable speed capability. The feed flow will be discharged into the first stage of the four stage RBC unit. Effluent from the RBC unit will flow by gravity to the 1.6 m secondary clarifier. Effluent from the secondary clarifier will flow by gravity along the ground and discharge into a manhole at the entrance of the building. Wasting will be batched from the bottom of the secondary clarifier by a pump to a small temporary holding tank. The holding tank will be used to determine the sludge production and quality. An overflow from the holding tank will discharge to the same manhole as the effluent from the RBC.

3.0 BIOLOGICAL AERATED FILTER (BAF) PROCESS

3.1 General

The BAF (Biocarbane) process was developed in Europe, principally France, in the late 1970s. The process is much like granular media water filter. Flow of primary treated effluent is downward through the immersed media. Air or oxygen is injected at the base of the filter to supply the oxygen uptake of the biofilm which attaches to the media. No secondary clarifier is required yielding one step aeration and suspended solids removal. Periodic backwash is required to control biofilm development, thus necessitating backwash water storage as in water filtration.



EXPECTED PILOT PLANT FEED RATE = 65 m3/d (12 US gpm)

Figure 3 - SCHEMATIC LAYOUT OF RBC PILOT PLANT

Table 2
RBC PILOT PLANT LOADING CHARACTERISTICS

Parameter	Pilot Plant	Typical Guideline
1. General		
• Flow (m^3/d)		
Avg.	65 (12 U.S. gpm)	
Peak	136 (25 U.S. gpm)	
• Influent Conc. (mg/L)		
$TBOD_5$	36	
TSS	26	
TKN	14.3	
2. RBC	2 m RBC, A_{surf} = 734 m^2	
• Organic Load ($gBOD_5/m^2 \cdot d$)	3.2 (0.65 lb $BOD_5/1000\ ft^2 \cdot d$)	12 - 20 g sol$BOD_5/m^2 \cdot d$ (2 - 4 lbs sol BOD_5/1000 ft•d)
• Ammonia Load ($gTKN/m^2 \cdot d$)	1.3 (0.26 lb TKN/1000 $ft^2 \cdot d$)	0.3 - 1.1 kg fil TKN/1000 $m^2 \cdot d$
• Hydraulic Loading	88	40 - 244 $L/m^2 \cdot d$ (90 $L/m^2 \cdot d$-EC)
3. Clarifier	1.6 m, A_{surf} = 2.0 m^3	
• SOR ($m^3/m \cdot d$)	32.5	

As a monthly effluent TP criteria of 0.5 mg/L may be imposed on the West Windsor PCP, effluent filtrations may be essential. The provision of biological treatment and suspended solids removal in one step may prove to be an attractive capital cost savings feature of the BAF treatment process.

Two municipal facilities which employ the BAF process are in or near operation in Quebec. At the Canton of Orford, a 2,200 m^3/day facility in operation since 1986, flow is first treated with physical/chemical precipitation with settling in a lamella clarifier. Clarified effluent is then discharged to the BAF process (Biocarbane type) for biological treatment and suspended solids removal. The Canton of Orford is required to meet a 0.5 mg/L TP criteria. A similar 132,000 m^3/day treatment system is under construction at Sherbrooke with anticipated start-up in 1990.

With the physically/chemically treated effluent quality obtained from the West Windsor PCP, it is anticipated that a high degree of nitrification could be achieved in the BAF process. Estimated nitrogen loadings for nitrification are 0.8 kg $NH_3\text{-}N/m^3 \cdot d$.

3.2 Pilot Plant Requirements

A flow schematic of the BAF process is shown in Figure 4. The BAF (Biocarbane filter) pilot plant will be supplied by John Meunier Inc., at a cost of $9,000 for a period not to exceed ten (10) months. The biocarbone filter unit contains a 2.5 m depth of filter media supported on a false floor with strainers for air and water backwash. The filter surface area is 0.5 m^2.

Typical design parameters of a BAF are presented in Table 3. The hydraulic and volumetric loading are the key design parameters of the BAF process which determine the feed flowrate. Limited by the ammonia loading rate, the feed pump (variable speed) will supply the BAF pilot plant from the common well at an average flowrate of approximately 45 m^3/d (8 U.S. gpm).

As shown in the process schematic of the pilot unit, the pilot plant will be supplied complete including the backwash water pump and the control panel for automatic operation of the backwash cycle. The following items are not included as part of the pilot plant as supplied by John Meunier Inc.:

- Feed Pump
- Process and Backwash Air
- Piping and Wiring
- Backwash Water Reservoir

For the BAF pilot plant, a backwash water reservoir of approximately 800 gal is required. Filtered water processed by the BAF unit is typically used to supply the backwash water reservoir, with the overflow to the effluent manhole. Automatic

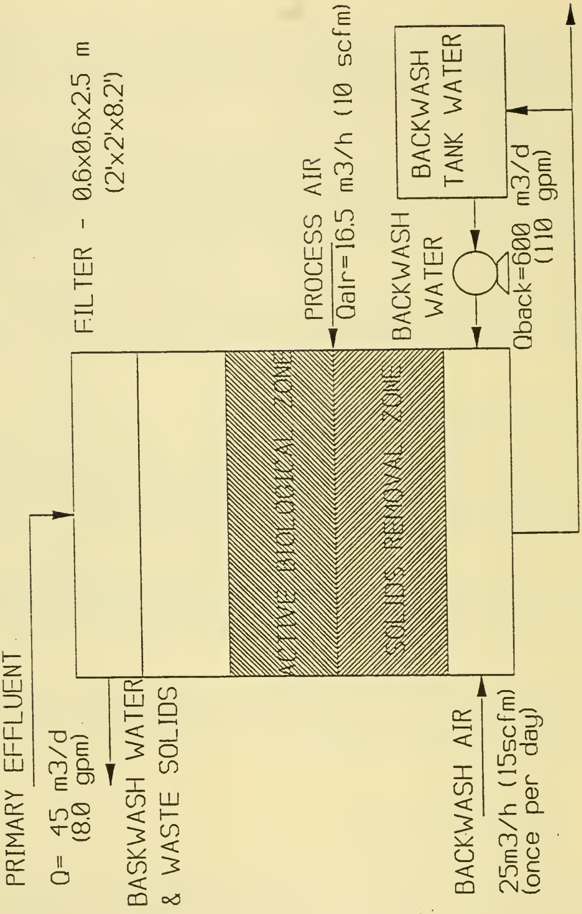


Figure 4 - SCHEMATIC LAYOUT OF BAF PILOT PLANT

Table 3
BAF (BIOCARBONE FILTER) LOADNG CHARACHTERISTICS

Parameter	Pilot Plant	Guidelines
1. General		
• Flow (m^3/d)		
Avg.	45	
Peak		
• Avg. Influent Conc. (mg/L)		
$TBOD_5$	36	
TSS	26	
NH_3	14.3	
2. BAF		
Surface Area	0.5 m^2	
Filter Volume	1.25 m^3	
Hydraulic Loading ($m^3/m^2 \bullet h$)	3.7	2.0 - 7.0 $m^3/m^2 \bullet h$
Organic Loading (kg $BOD_5/m^3 \bullet d$)	1.3	2.0 - 6.0 kg $BOD_5/m^3 \bullet d$
TKN Loading ($kg/m^3 \bullet d$)	0.7	<0.7 kg $TKN/m^3 \bullet d$
HRT (min)	40	30 - 60 min

controls and instrumentation operate the backwash air and water cycles, which takes place once per day during normal operation of the unit.

The process and backwash air requirements for the BAF pilot plant will be provided by Meunier. Instrumentation air is also required for the operation of the backwash cycle and process control valves which are supplied with the pilot plants.

4.0 "NON-TRADITIONAL" ACTIVATED SLUDGE

4.1 General

The activated sludge process is the most common secondary treatment process used for the treatment of municipal sewage in Canada. The City is very familiar with this type of treatment from the operation of the Little River PCP.

The activated sludge process involves two main components, these being the aeration tank and the secondary clarifier. The aeration tank provides for the biological treatment of the organics and can provide nitrification. The secondary clarifier settles and concentrates the solids for recycle and wasting.

The results of the process evaluation of the Little River PCP indicates that a "non-traditional" activated sludge plant should be able to reduce the remaining organic load and provide nitrification at West Windsor. The Little River PCP operated at an average hydraulic retention time of 2.5 h, and an solids retention time of 7 d. The secondary clarifier also was able to operate at peak flows in excess of MOE guidelines.

In addition to operating the pilot facility at conditions similar to the Little River PCP, a provision has been made to operate an anoxic zone upstream of the aeration tank. The main purpose of the anoxic zone will not be for biological nutrient removal but for settleability control.

An anoxic environment is generally defined as an environment where molecular oxygen is absent but oxidized nitrogen (nitrite, nitrate) is present. The treatment objective is to achieve a high degree of soluble substrate removal in the anoxic zone. The anoxic zone provides a short contact (0.33 h) time and initial high F/M ratio immediately following the primary clarifier. Nitrate is supplied by both the return sludge and a nitrate-rich mixed liquor recycle from the aeration basin. Within the anoxic zone, floc-formers may use nitrate as a electron acceptor in the absence of molecular oxygen. This allows these organisms to exist in an exogenous phase in the high F/M anoxic region. Since filamentous organisms are obligate aerobes, they will not become active until the subsequent aeration step. At the aeration step most of the readily degradable, soluble organic have already been removed from the wastewater. This prevents the filamentous organism population from far exceeding a balanced limit and causing a sludge bulking problem. Conditions favour the formation of floc-formers in the system at the expense of filamentous organisms.

4.2 Pilot Plant Requirements

The "non-traditional" process will be operated as a bench-scale plant, a schematic of the process is shown in Figure 5. The bench-scale plant will be provided by CH2M HILL.

The pilot plant will be loaded at an average flow of approximately 70 mL/min. The resulting design guidelines are reviewed in Table 4. The results show that the pilot plant will be operated at loading parameters which exceed the MOE guidelines, but are similar to the values the Little River PCP was operated at during our process evaluation.

The bench-scale unit will be located in the composite dryer building and will be fed primary effluent from the common feed well. Flow will proceed through the anoxic zone followed by the aeration tank and the secondary clarifier. Return solids will be pumped to the anoxic zone. Wasting will be manually controlled.

Table 4
ACTIVATED SLUDGE PILOT PLANT LOADING CHARACTERISTICS

Parameter	Pilot Plant	Guidelines
1. General		
• Flow (m^3/d)		
Avg.	0.1 (70 mL/min)	
Peak	0.2 (140 mL/min)	
• Influent Conc. (mg/L)		
$TBOD_5$	36	
TSS	26	
NH_3	14.3	
2. Anoxic Zone	V = 1.4L	
• HRT (h)	0.33	
3. Aeration Tank	V = 10.5 L	
• HRT (h)	2.5	6 h
• Organic loading ($kg\ BOD_5/m^3 \cdot d$)	0.34	0.3 - 0.71 $kg\ BOD_5/m^3 \cdot d$
• F/M (d^{-1})	0.18	
4. Secondary Clarifier	$A_s = 266\ cm^2$	
Peak SOR ($m^3/m^2 \cdot d$)	7.5	29.4 (nitr) $m^3/m^2 \cdot d$

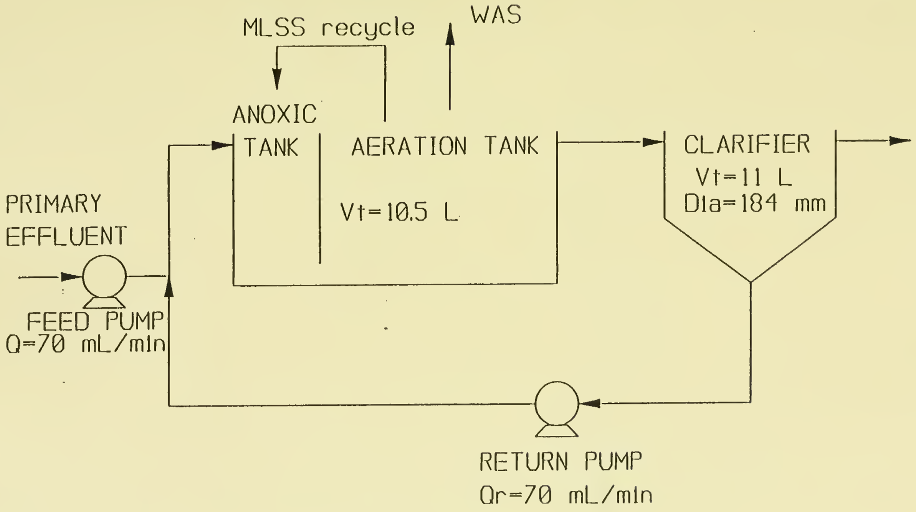


Figure 5 - SCHEMATIC LAYOUT OF ACTIVATED SLUDGE PILOT PLANT

Appendix C

PILOT PLANT MONITORING SCHEDULE

PILOT PLANT DAILY MONITORING SHEET

DATE: ____________
AIR TEMPERATURE: ____________
WEATHER: ____________

PARAMETER	TF/SC	RBC	BAF	ACT SLUDGE
FLOWS (m^3/h)				
INFLUENT				
RAS		NA	NA	
WAS				
SOLIDS (mg/L)				
Primary Eff.				
Final Effluent				
MLSS (MLVSS)		NA NA	NA NA	
RAS		NA	NA	
WAS				
OPERATIONS				
INF. T° (°C)				
EFF. T°				
PROCESS T°				
CLAR SLUDGE BLK (m)			NA	
pH				
SETTLED VOL (%)		NA	NA	
SVI		NA	NA	
DO CONC.				
OTHER				

COMMENT: ________________________________
____

NA - indicates not applicable

Appendix D

DISCUSSION OF EFFLUENT REQUIREMENTS (SEPTEMBER 25, 1990 MEETING NOTES)

Please contact Tony Ho (416-314-3877) for information/data contained in this Appendix.

Appendix H

FULL-SCALE SITE INSPECTION SUMMARIES

MEMORANDUM

CH2M HILL

CITY OF WINDSOR
WEST WINDSOR PCP ALTERNATIVES EVALUATION

DATE: June 28, 1990

SUBJECT: Summary of Site Visit to two BAF (Biocarbon Facilities in Quebec

PROJECT: ONT29343.A0

ATTENDEES:
Mr. L. Romano, City of Windsor
Mr. K. Woods, City of Windsor
Mr. M. Maki, City of Windsor
Mr. G. Meunier, Meunier Inc.
Mr. F. Seguin, Meunier Inc.
Mr. G. Daigger, CH2M HILL
Mr. J. Stephenson, CH2M HILL
Mr. M. Newbigging, CH2M HILL

Site visits to two BAF (biological aerated filter) units located in Quebec were conducted on May 8, 1990. Representatives from the City of Windsor and CH2M Hill were taken on the site visits by Mr. G. Meunier and Mr. F. Seguin of John Meunier Inc. Both plants are designed for organic or BOD_5 removal only. The plants visited were:

Sherbrooke, Quebec

The plant is currently under construction and is expected to be completed by next year. The plant average daily capacity is 16 MIGD (72,000 m^3/d) with a peak daily capacity of 29 MIGD (132,000 m^3/d). The plant will provide primary clarification followed by the BAF units. The backwashed solids will be thickened in a filter press.

The secondary treatment will be accomplished in fourteen separate BAF tanks. The BAF units are housed inside a building, and will be enclosed in individual fibreglass domes when complete. The separate enclosures are to protect the building interior from corrosion. Each tank was approximately 4 m long by 2.5 m wide and 4 m deep. The media is supported on a false floor which is raised about 1 m off the tank bottom. Backwash air and effluent water will be fed through the false floor. Process air will be supplied to a series of perforated tubes laid out on top of the false floor. Each BAF unit will be taken out of service in series; backwashing will occur in five cycles (backwash time approx. 10 min) starting with backwash air and followed by backwash water. The water level in the tank will rise and a siphon will take the solids laden backwash water to storage prior to being thickened in a filter press. Clean backwash water is to be stored in a tank prior to use in backwashing.

Although the plant will not be operational for sometime, the visit gave a good indication of the workings of a large scale facility. Some interesting aspects of the facility included:

- the false floor and feeder system for the backwash air and water
- a possible layout of the plant, showing the orientation of the channels, piping and tanks
- the relative scale of a plant, which will be approximately half the size of the West Windsor PCP

Canton d'Orford, Quebec

The Orford plant is a 0.5 MIGD (2200 m^3/d) facility used to treat domestic sewage. The plant has been in operation for 3 years and services a ski hill resort community. As a result the flows were very low on the day of our visit. The plant has influent screening, followed by a primary, lamella type clarifier and two BAF units. The operational staff initiated a backwash cycle and we were able to observe the process.

Primary effluent flows by gravity to the two BAF (Biocarbon) units. Treated effluent is taken off the bottom and flows through an effluent backwash tank. During the backwash cycle, feed flow is shut off and a series of backwash cycles occur (air and effluent water). A separate channel is used to take the solids laden backwash water to an aerobic digester prior to thickening. Separate but similar sized blowers provide the process and backwash air. Backwashing is conducted on one tank at a time.

During the visit, G. Meunier indicated that the City can expect the pilot BAF or Biocarbon unit to be onsite around the first week in June. Meunier will provide additional information to the City of Windsor in terms of piping, wiring and control valves.

Meunier also indicated that there is a another plant, Boisbriand with a capacity 5.5 MIGD (25,000 m^3/d), located near Montreal which could also be viewed in the future. The plant has been in operation for less than a year.

M E M O R A N D U M CH2M HILL

CITY OF WINDSOR
WEST WINDSOR PCP ALTERNATIVES EVALUATION

DATE: June 28, 1990

SUBJECT: Site Visit to Investigate the Trickling Filter/Solids Contact Process

PROJECT: ONT29343.A0

ATTENDEES:
Mr. L. Romano, City of Windsor
Mr. K. Woods, City of Windsor
Mr. M. Maki, City of Windsor
Mr. D. Duklow, City of Windsor
Mr. C. Cowles, Cowles Equipment Sales Inc.
Mr. J. Stephenson, CH2M HILL
Mr. P. Robson, CH2M HILL
Mr. M. Newbigging, CH2M HILL

On June 1, 1990 representatives of the City and CH2M Hill were taken on a tour of two trickling filter facilities by Curt Cowles, sales representative for Munters Corp. The two sites were Wauseon, Ohio and Coldwater, Michigan. The following is a brief description of the visits to each plant.

Wauseon, Ohio

The Wauseon, Ohio plant began operation in February 1989. The plant capacity is 1.5 (US)mgd average and 3.0 (US) mgd peak facility. The plant is currently treating a little over 1 mgd. The plant's effluent quality permit calls for the plant to obtain 15, 17 and 1 mg/L for Carbonaceous BOD_5, total suspended solids and total phosphorous. Although the City does not appear to have an effluent ammonia requirement, the plant has been nitrifying since three months after startup. Data reviewed for the plant is attached and has been updated with data sent by the plant following the visit. The City's sewer system is a partially combined collection system. The influent to the plant is conveyed by two sewer mains, one a 30-inch and the other an 18-inch diameter pipe.

The plant is a trickling filter/ solids contact process with fine raw screening, primary clarification and disinfection by chlorination/dechlorination. The waste activated sludge is cothickened in the primaries and pumped with the primary effluent sludge to an anaerobic digester followed by two secondary digesters or holding tanks. Digested sludge is spread on farm fields as a soil conditioner.

Some of the points discussed and observed concerning the trickling filters/solids contact process were:

- The trickling filters are open at the top and are contained in two concrete tanks with a central access.
- Ventilation is provided around the base of the filter by approximately eight vents each approximately 2 by 4 ft.
- The media is supported on a number of concrete columns about 4 inches in diameter. The base of the trickling filter is sloped to a effluent well.
- The concrete tanks are not lined, although Mr. C. Cowles indicates it is advisable to do so. Unlined concrete tank tend to crack and deteriorate due to freeze/thaw problems.
- The plant pumps primary effluent to the trickling filters via centrifugal pumps.
- The feeder system has four arms and was operating at one revolution per minute, wetting the surface every 15 seconds. The feeder arm was operated by the pressure of the feed flow and is therefore variable. The discharge nozzles consisted of 1 to 1.5 inch openings with a hinged flap which tended to spray the flow onto the media. The nozzles were unevenly spaced with a total of about ten nozzles on each arm.
- The superintendent, Tim Hausch, indicated that the plant recycles about 100 percent of the trickling filter effluent around the filter. This seems reasonable since the design wetting rate is only 0.12 US gpm/ft^3, therefore a 100 percent recycle results in a wetting rate of 0.24 US gpm/ft^3. The primary effluent loading at this plant is about twice that of West Windsor, therefore a 100 percent dilution of the trickling filter feed results in a feed similar to West Windsor.
- The trickling filter effluent flows by gravity to two solids contact tanks which have fine bubble diffusers. The dissolved oxygen levels in the tanks was 4 to 6 mg/L according to the superintendent; however routine measurements are not made.
- The two secondary clarifiers were 16 feet deep with flocculator centre wells with slow mixing.

One of the interesting aspects of the plant was the close proximity of the plant to a neighbouring housing development. The superintendent indicated that they have good relations with the community and have had no complaints thus far. The superintendent agreed to send the City monthly summaries for the plant from October 1989 to present. We thanked Mr. Tim Hausch for the tour and he indicated that if we had any further questions he would be happy to respond to them. The plant's phone number is (419) 335-3026.

Coldwater, Michigan

The plant services the town of Coldwater, population approximately 9,500. The current capacity of the plant is 2.7 mgd, and the plant currently treats about 2.1 mgd. The town has a high industrial component, including a rendering plant which discharges about 125,000 gal/d of high strength waste to the plant. The plant was originally built in 1951 and consisted of primary clarification, rock media trickling filters and secondary clarifiers. The plant has been upgraded in 1972 and in 1988 to its current configuration, which includes:

- grit removal
- rectangular primary clarifiers
- three 115 foot diameter, 8 foot deep covered plastic media trickling filters
- solids contact aeration tank with fine bubble diffusers (Sanitaire)
- two secondary clarifiers
- pressure filtration
- chlorination/dechlorination
- anaerobic digestion

Our group was shown the plant by Mr. Greg Rubley. The superintendent is Mr. Dave McKay. The plant's current summer discharge requirements are 20, 15 and 0.47 mg/L for BOD_5, total suspended solids and total phosphorus. The plant has no nitrification requirement and does not sample the effluent ammonia. Observation of the effluent data indicates the plant obtains good BOD_5 and solids removal and excellent total phosphorus removal. The total phosphorus is generally less than 0.5 mg/L prior to the pressure filters. It may be possible that the trickling filter has had some impact on the total phosphorus removal. In terms of the trickling filter/solids contact process, the following observations were made.

- The plant has covered its three trickling filters with aluminum domes, mainly due to freezing problems during winter months. The walls of the old rock filters, which are shown in pictures at the plant, rose only slightly above the media and therefore the feeder system was exposed to the wind. The new filter feeder system is about eight feet below the top of the concrete containers. This would probably shelter the feeder system from the wind.

- The feeder system operated at about two revolutions per minute, during our tour. Wetting therefore occurred every 7.5 seconds.

- The nozzles were slightly different from the ones seen at the Wauseon WWTP. The nozzles were on the sides of the feeder pipe, and had about 1.0 inch openings. The flow exited the orifice and distributed by a splash plate shaped like an arrowhead.

The splash plate was attached to the nozzle and was bent over the opening in the pipe causing the flow to be splayed over the media.

- Greg Rubley indicated that the plant is currently only operating two of the filters to retain some load for the aeration tanks.

- The filters had no ventilation, other than through the top of the media. Curt Cowles indicated this is common for shallow filters which are less than 10 feet deep.

- The aeration tank had fine bubble diffusers, the units are Sanitaire ceramic disks. Similar disks are installed at the Little River PCP plant no. 2. Greg Rubley indicated that the plant uses the Sanitaire cleaning system about twice a year.

- A brief review of the current weeks data indicated the plant was obtaining 99, 98 and 93 percent removal for BOD_5, total suspended solids and total phosphorus. The approximate concentrations are shown in the table below.

APPROXIMATE CONCENTRATIONS OF COMPLIANCE PARAMETERS

Parameter	**Concentration mg/L**			
	Raw	**TF Effluent**	**Sec. Effluent**	**Filter Effluent**
BOD_5	200	80	10	2
TSS	150	30	6	3
TP	2.3	--	0.15	0.15

The Coldwater plant's phone number is (517) 279-9531.

WEST WINDSOR PCP - PLANT UPGRADE
ONT29343.A0
SUMMARY OF TF/SC PROCESS DATA

PLANT : WAUSEON WWTP, OHIO
TF SIZE : TWO 75FT DIA X 14FT DEEP

DATE	FLOW	INFLUENT (mg/L)			PRIMARY EFFLUENT (mg/L)			FINAL EFFLUENT (mg/L)		
	(mgd)	CBOD5	TSS	NH3	CBOD5	TSS	NH3	CBOD5	TSS	NH3
Jan-89	1.18	124	138	10.1				71	62	8.8
Feb-89	0.99	150	153	15.6				74	60	12.5
Mar-89	1.17	136	239					37	44	12.1
Apr-89	1.38	112	113	10.3	90	54	11.2	9	12	11.7
May-89	1.25	172	171	15.9	122	73	15.2	6	12	6.9
Jun-89	1.54	108	175	12.0	87	81	10.2	4	8	0.9
Jul-89	0.90	151	247	18.8	116	90	18.0	3	5	0.3
Aug-89	0.85	162	363	19.6	125	127	16.8	4	6	0.2
Sep-89	1.04	132	334	15.6	104	101	11.9	3	5	0.1
Oct-89	0.72	189	283	27.5	153	123	23.6	3	7	0.7
Nov-89	0.76	187	186	25.9	147	94	20.5	3	10	0.7
Dec-89	0.55	202	178	28.8	163	96	27.2	10	21	7.7
Jan-90	1.38	108	159	11.8	102	80	10.5	4	9	1.3
Feb-90	1.91	86	124	8.1	80	83	7.5	5	9	2.0
Mar-90	1.66	99	109	9.2	93	66	8.5	3	5	0.8
Apr-90	1.45	117	165	8.9	91	76	7.5	3	9	0.1
AVG =	1.17	140	196	15.9	113	88	14.5	15	18	4.2
STD =	0.36	34	74	6.7	26	20	6.1	23	19	4.7
MIN =	0.55	86	109	8.1	80	54	7.5	3	5	0.1
MAX =	1.91	202	363	28.8	163	127	27.2	74	62	12.5

CITY OF WINDSOR
WEST WINDSOR PCP ALTERNATIVES EVALUATION

DATE: June 29, 1990

SUBJECT: Summary of Site Visit to Two RBC Facilities

PROJECT: ONT29343.A0

ATTENDEES:
Mr. K. Woods, City of Windsor
Mr. M. Maki, City of Windsor
Dr. R. Dryman, City of Windsor
Mr. H. Moneith, MOE
Mr. J. Stephenson, CH2M HILL
Mr. P. Robson, CH2M HILL
Mr. M. Newbigging, CH2M HILL

Site visits to two RBC facilities were conducted on June 28, 1990. The two sites visited were the Niagara Falls and Guelph WPCPs. A summary schematic of each plant and its 1988 MOE analytical data are attached for background information. A brief description of each visit follows.

Niagara Falls WPCP

Our group was toured around the plant by Mr. M. Lywood, Mr. J. Pine, and Mr. J. Adams of the Regional Municipality of Niagara. The Niagara Falls (Stamford) WPCP was originally built in 1963 to provide primary treatment for an average flow of 45,500 m^3/d. In 1978, the primary capacity was expanded to 54,600 m^3/d. In 1986, the capacity was increased to 68,200 m^3/d of secondary capacity with the addition of RBCs and final clarifiers.

The secondary treatment is provided in 35 rotating biological contactors (RBC) in five trains of seven stage units. The RBCs provide approximately 345,000 m^2 of media area for biomass growth. The RBC effluent flows to four 33.5 m diameter final clarifiers. The main discussion and observations during the tour were:

- The Region has had no problems with any overgrowth leading to shaft failures. The RBC units however do have the provision to reverse the motors to s~~luff-off~~ slough off excess growth.

- The units operate at about 1.5 rpm. Each RBC is powered by a 5 hp motor. The plant has the provision of operating a train of RBCs on backup power to avoid excess growth during long power outages.

- The RBCs have been operating at the Niagara Falls plant for the last three and a half years with no major problems encountered.

- The units are covered with fibreglass enclosures with a large opening at the motor drive and smaller accesses for viewing the media and biomass growth.

- Biomass growth was heavy on the first few stages, the growth was thick and grey in colour. The growth was significantly reduced by the last stage. The growth on the extremely opposite train was less, probably indicating an imbalance in the loading.

- Each stage has baffles between subsequent stages to avoid short circuiting. In addition, each train can be taken out of service separately.

- The plant aerates the channel upstream and downstream of their RBCs to maintain the solids in suspension. This air may also be used as an oxygen source for the RBC biomass.

- The plant adds ferric chloride for phosphorus removal in the RBC effluent.

- Waste activated solids are pumped from the secondary clarifiers by six diaphragm pumps on timers. The plant indicate they try to maintain no blanket since all of the sludge is wasted. Primary and waste activated sludge are pumped separately to the digester.

- The plant is designed to provide conventional secondary treatment with C of A requirements of 25, 25, and 1 mg/L for total BOD_5, suspended solids and total phosphorus. The plant was not designed to nitrify, but does obtain partial nitrification occasionally. A brief look at historical data indicated that in February 1990, partial nitrification was obtained. The analytical data for this month are shown below. However, the next monthly sample indicated negligible nitrification.

We thanked the Region for the tour and they indicated that should we have any further questions to call them at (416) 354-2754 or FAX at (416) 374-7188.

ANALYTICAL DATA FOR FEBRUARY 1990		
Parameter	**Raw**	**Effluent**
$TBOD_5$	82	19.5
TSS	93.7	17.6
TO	3.6	0.7
TKN	25	12

Guelph WPCP

Our group was taken around the Guelph RBC units by Mr. J. Sanvido, Manager of Water and Wastewater Services. We mainly focused on the RBC and effluent filters at the plant.

The plant's hydraulic design capacity is 54,500 m^3/d and the plant treated 41,700 m^3/d in 1988. The plant provides screening, grit removal, primary clarification, conventional activated sludge removal, nitrifying RBCs, effluent filtration and chlorination.

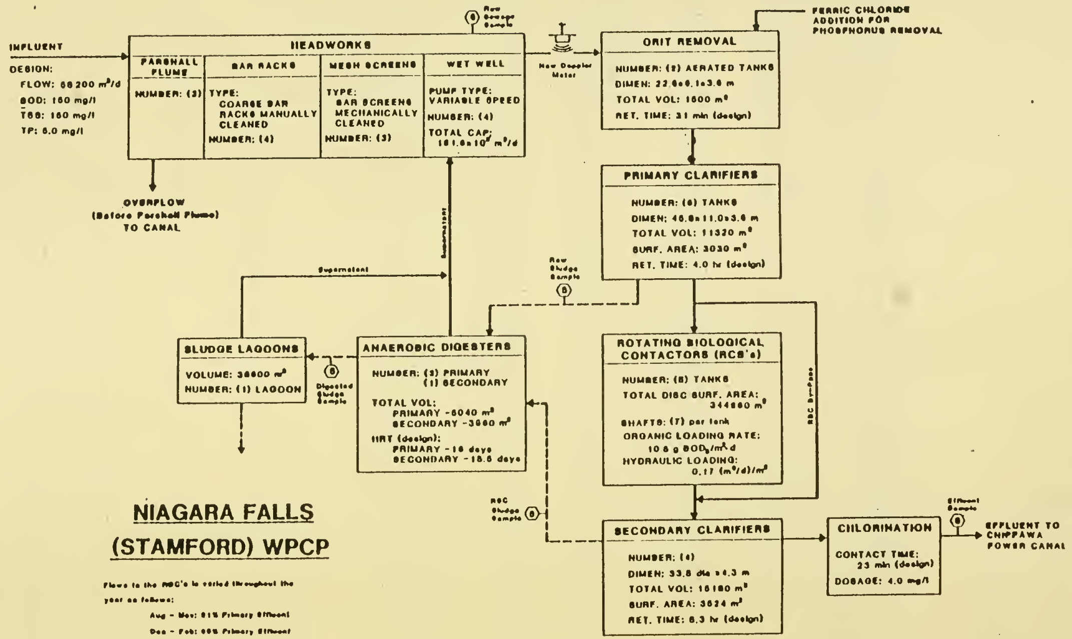
Based on plant experience the RBCs receive influent ammonia concentrations from 12 to 20 mg/L and generally provide full to partial nitrification with effluent concentrations from 0.1 to 2 mg/L. Occasionally in the winter the effluent ammonia concentrations rise to 4 - 4.5 mg/L. The plant's effluent requirements calls for seasonal nitrification.

The following items were discussed or observed:

- The plant is not effected by most rainfall or snow melt events since the entire sewer system is separate. This differs from Windsor which receives a significant infiltration/inflow component.
- The plant has 32 RBC units, in four, eight stage trains. The plant has 400,000 m^2 of media surface area.
- The RBCs are air driven by three blowers located in the RBC influent lift station. The RBCs have vents on the outer edge of each RBC into which air is forced to provide rotation. The RBCs rotate at about 1 rpm. The system is easier than the motor driven in terms of lesser number of motors, for example, three blowers instead of 32 small motors. Also three blowers provides backup for blower failures whereas there is no backup on direct drive units. However,

the biomass on these units was very slight, since it is only for nitrification, therefore the ability to drive a combined organic and nitrifying RBC by air alone is unknown.

- Mr. Sanvido indicated the City have had no problems with the RBCs, and have found them to be very easy to operate and a low maintenance item. The plant checks the rotation speed every three weeks and greases the units once every three months.
- As mentioned the biomass growth on these RBCs is very light. The first stage has a light slime film about a millimetre or two thick. The last stages film is very light, the edge disk had almost no slime.
- After effluent filtration the plant obtains excellent effluent averaging 4,2 and less than 0.5 mg/L of BOD_5, total suspended solids, and total phosphorus.



NIAGARA FALLS (STAMFORD) WPCP

Flows to the RBC's is varied throughout the year as follows:

Aug - Nov: 81% Primary Effluent
Dec - Feb: 66% Primary Effluent
Mar - Jul: 100% Primary Effluent

CANVIRO

MUNICIPALITY : NIAGARA FALLS
PLANT : STAMFORD WPCP
WORKS NUMBER : 120001363
TREATMENT : ROTATING BIOLOGICAL CONTACTOR
: PHOSPHORUS REMOVAL-CONTINUOUS
:
DESIGN CAPACITY : 58.10 (1000 M3)

REGION : WEST CENTRAL
DISTRICT : NIAGARA, REG. MUN.
OPERATING AUTHORITY : MUNICIPAL
WATERCOURSE : CHIPPAWA POWER CANAL
MINOR BASIN : ONTARIO
MAJOR BASIN : GREAT LAKES
POPULATION SERVED : 67,835

	JAN	FEB	MAR	APR	MAY	JUN	JUL	AUG	SEP	OCT	NOV	DEC	ANNUAL AVERAGE	NUMBER EXCEED
AVG. DAILY FLOW (1000 M3)	44.62	56.56	58.88	57.13	53.58	56.07	58.36	61.37	83.80	59.90	57.30	54.40	58.50	
BOD5														
INFLUENT (MG/L)	92.0	71.0	55.0	72.0	66.0	28.0	81.0	97.0	83.0	125.0	74.0	81.0	77.1	
EFFLUENT (MG/L)	14.0	21.0	8.9	15.0	14.0	15.6	12.0	16.0	23.0	19.0	17.7	14.7	15.9	
CONCENTRATION LIMIT (MG/L)													25.0	
LOADING (KG/D)	624.68	1187.76	524.03	856.95	750.12	874.69	700.32	981.92	1927.40	1138.10	1014.21	799.68	930.15	
SUSPENDED SOLIDS														
INFLUENT (MG/L)	126.0	110.0	99.0	106.0	123.0	134.0	102.0	122.0	141.0	137.0	122.0	109.0	119.3	
EFFLUENT (MG/L)	21.0	21.0	17.8	19.3	18.0	18.0	17.0	21.0	14.0	23.0	22.5	21.4	19.5	
CONCENTRATION LIMIT (MG/L)													25.0	
LOADING (KG/D)	937.02	1187.76	1048.06	1102.60	964.44	1009.26	992.12	1288.77	1173.20	1377.70	1289.25	1164.16	1140.75	
TOTAL PHOSPHOROUS														
INFLUENT (MG/L)	4.2	3.4	2.7	3.3	3.7	3.3	3.9	3.0	3.3	2.8	3.2	2.8	3.4	
EFFLUENT (MG/L)	0.8	0.9	0.4	1.0	0.7	0.6	0.8	0.9	0.6	0.5	0.8	0.3	0.7	
CONCENTRATION LIMIT (MG/L)	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0		
LOADING (KG/D)	35.69	50.90	23.55	57.13	37.50	33.64	46.68	55.23	50.28	29.95	45.84	16.32	40.95	

SUMMARY

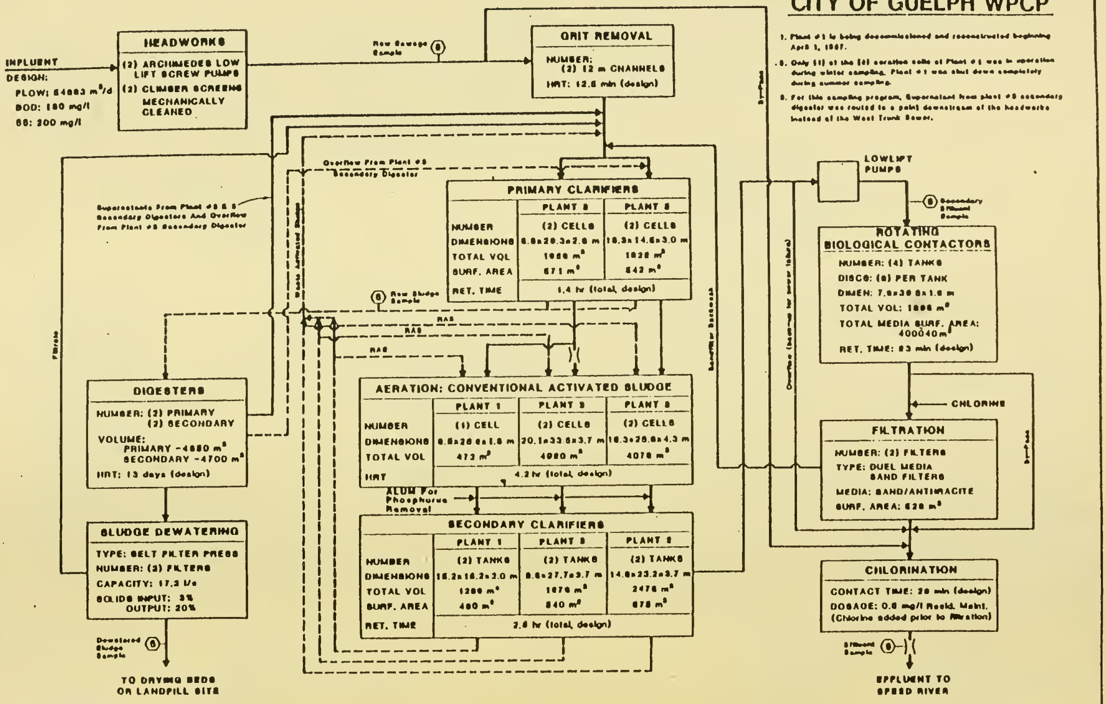
PARM	CRITERIA	COMPLIES WITH CONC
BOD	25.0 MG/L	YES
SS	25.0 MG/L	YES
TP	1.0 MG/L	YES

NOTE:
BOD - HAS NO CRITERIA
SS - HAS NO CRITERIA
TP - HAS NO CRITERIA
* - EXCEEDS EFFLUENT CRITERIA
SPACE (EG. " ") INDICATES A VALUE OF ZERO
ND - NO DATA
NA - NOT APPLICABLE

NO REMEDIAL MEASURES REPORTED

CITY OF GUELPH WPCP

1. Plant #1 is being decommissioned and reconstructed beginning April 1, 1987.
2. Only (1) of the (2) aeration cells at Plant #1 was in operation during winter sampling. Plant #1 was shut down completely during summer sampling.
3. For this sampling program, Supernatant from plant #3 secondary digester was routed to a point downstream of the headworks instead of the West Trunk Sewer.



MUNICIPALITY : GUELPH
PLANT : GUELPH WPCP
WORKS NUMBER : 120003094
TREATMENT : CONVENTIONAL ACTIVATED SLUDGE
: PHOSPHOROUS REMOVAL-CONTINUOUS
: EFFLUENT POLISHING
DESIGN CAPACITY : 54.55 (1000 M3)

REGION : WEST CENTRAL
DISTRICT : WELLINGTON
OPERATING AUTHORITY : MUNICIPAL
WATERCOURSE : SPEED RIVER
MINOR BASIN : ERIE
MAJOR BASIN : GREAT LAKES
POPULATION SERVED : 82,000

	JAN	FEB	MAR	APR	MAY	JUN	JUL	AUG	SEP	OCT	NOV	DEC	ANNUAL AVERAGE	NUMBER EXCEED
AVG. DAILY FLOW (1000 M3)	41.14	42.34	43.12	43.44	42.05	41.04	36.56	40.61	43.08	41.12	44.33	41.17	41.73	
BOD5														
INFLUENT (MG/L)	183.0	236.0	ND	160.0	166.0	166.0	212.0	164.0	176.0	190.0	163.0	179.0	181.4	
EFFLUENT (MG/L)	11.7	13.0	ND	12.9	5.1	10.7	7.4	8.2	5.0	4.6	7.5	4.0	8.2	
CONCENTRATION LIMIT (MG/L)													25.0	
LOADING (KG/D)	481.33	550.42	ND	560.37	218.53	439.12	270.54	333.00	215.40	189.15	332.47	164.68	342.19	
SUSPENDED SOLIDS														
INFLUENT (MG/L)	210.0	108.0	ND	160.0	192.0	105.0	100.0	238.0	208.0	202.0	209.0	163.0	194.8	
EFFLUENT (MG/L)	13.3	14.0	ND	14.8	9.8	12.0	9.4	12.3	9.6	8.8	10.7	8.9	11.2	
CONCENTRATION LIMIT (MG/L)													25.0	
LOADING (KG/D)	547.16	592.76	ND	642.91	419.93	492.48	343.66	499.50	413.56	361.05	474.33	366.41	467.38	
TOTAL PHOSPHOROUS														
INFLUENT (MG/L)	5.3	5.4	ND	5.3	6.9	6.3	6.5	7.9	11.4	6.4	6.9	7.4	6.9	
EFFLUENT (MG/L)	0.5	0.7	ND	0.6	0.5	0.5	0.4	0.5	0.5	0.5	0.9	0.9	0.6	
CONCENTRATION LIMIT (MG/L)	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0		
LOADING (KG/D)	20.57	29.63	ND	26.06	21.42	20.52	14.62	20.30	21.54	20.56	39.89	37.05	25.04	

SUMMARY

PARM	CRITERIA	COMPLIES WITH CONC
BOD	25.0 MG/L	YES
SS	25.0 MG/L	YES
TP	1.0 MG/L	YES

NO REMEDIAL MEASURES REPORTED

NOTE:
BOD - ASSESSED ANNUALLY
SS - ASSESSED ANNUALLY
TP - ASSESSED MONTHLY
* - EXCEEDS EFFLUENT CRITERIA
SPACE (EG. " ") INDICATES A VALUE OF ZERO
ND - NO DATA
NA - NOT APPLICABLE

CH2M HILL
ENGINEERING
LTD.

CITY OF WINDSOR DEPARTMENT OF PUBLIC WORKS

EVALUATION OF INNOVATIVE ALTERNATIVES FOR THE UPGRADING OF THE WEST WINDSOR POLLUTION CONTROL PLANT

SITE VISIT SUMMARY

DATE: April 2 to 6, 1991

PROJECT: ONT29343.D0

ATTENDEES:

Mr. L. Romano, City of Windsor
Mr. H. Horneck, LCBA
Mr. J. Stephenson, CH2M HILL
Mr. M. Maki, City of Windsor
Dr. G. Daigger, CH2M HILL
Mr. M. Newbigging, CH2M HILL

SUBJECT: Trip Summary
Full Scale Trickling filter/Solids Contact Plant Site Visits in the Western United States

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To assist in understanding the Trickling Filter/Solids Contact (TF/SC) process and the implications of a possible full-scale plant at the City of Windsor's West Windsor PCP facility, a site inspection of six full-scale plants in the Western United States was conducted from April 2 to 6, 1991. A copy of the tour schedule and agenda is attached. A separate photographic volume is provided with photos for each plant.

The following is a summary of the tour and the main points raised and discussed with the various plants' operational staff. Additional design and process data from each plant is attached and form an integral part of this summary. These data are referenced in each plant's description below and should be reviewed with these narratives. The plant schematics and, in some cases, process design data are also critical to a clear understanding of each plant. Prior to using any of the data in this report for pre- or final design purposes, details should be reconfirmed by contact with the municipalities concerned.

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1) THE DUCK CREEK WPCP, CITY OF GARLAND, TEXAS

Date: 2 April 1991; 1:30 to 5:00 p.m.

Contacts: Mr. Lynn Norton, Director of Wastewater
Mr. Rick Watson, Lead Process Manager
Mr. Sid Richards, Plant Superintendent

Attachments: Duck Creek WPCP flow schematics and plant layouts; startup data and operational data from 1985 to 1991; plant 1990 Annual Report. These summaries are exceptional and should be reviewed with this narrative.

Address: City of Garland, Texas

Plant design: 30 USmgd; 50 million US gallon inline equalization; TF towers can hydraulically handle 40 USmgd; SC can hydraulically handle additional 20 USmgd; nitrification required.

Mr. Ron Seiger of CH2M HILL's Dallas office made local arrangements and accompanied our group to the plant. Ron introduced our group to the above City representatives. The City showed an introductory slide show to familiarize us with the plant and its layout. Mr. Rick Watson lead our group on a tour of the plant and provided background and addressed our questions. The following main items and discussion points were made:

- This plant is a single stage nitrifying TF/SC facility. It was specifically designed to nitrify in a single stage.

- The plant has 20 cylindrical dual-media final effluent filters; the media are sand and anthracite. This provides a final polished effluent.

- The second phase of the plant construction was completed in June 1987, resulting in a capacity of 30 USmgd.

- The plant's effluent limits were set in May 1986 at 10 and 15 mg/L for BOD_5 and SS. In November, 1988 an effluent ammonia limitation of 5 mg/L winter and 2 mg/L summer was imposed on the plant. The TF/SC process was put online in May 1986.

- The plant has inline flow equalization for 50 million gallons U.S. The equalization tanks are large and are cleaned regularly. Therefore, the plant does not have to treat diurnal flow variations. Annual precipitation is 34 inches with some intense storm events (to 8 inches per hour). The large equalization capacity is to avoid pushing storm events through the plant. The City has 900 miles of sewer lines and 54,000 connections. A 15-mile forcemain feeds the plant. The Cities serviced do not have combined sewers. These conditions should be noted for Windsor circumstances.

- Most pumps at the facility were located outside, without enclosures. This should be noted carefully for Canadian climatic conditions in connection with cost comparison.

- Plant design was based on a maximum month and a peak two-hour flow rate of 2.5 times the average flow. Secondary clarifier overflow rates were established at a maximum value of 400 USgpd/ft^2 and a peak 2-hour value of 1000 USgpd/ft^2. However, these conditions are not observed due to equalization of the influent flow rate. The final clarifiers have a 14 ft. side water depth.

- Trickling filter towers (140 ft diam) are plastic media with rotary distributors and effluent recirculation at low flow conditions. Two towers are 14 ft. in depth and two are 16 ft. in depth for a total of four towers. Trickling filter units can withstand about 40 USmgd hydraulically. Trickling filter effluent is pumped to the SC unit. Hydraulically, an additional 20 USmgd of primary effluent can bypass the TF towers to the SC unit.

- One of the trickling filter arms was rotating at approximately one revolution per minute. Another was operating at one revolution every two minutes. The plant is investigating slowing down the arms, by opening holes on the side opposite to the rotating direction.

- The solids contact tank has three passes. The return solids are re-aerated in the first compartment, the trickling filter effluent enters the front of the second compartment. The solids contact tank has fine bubble (Sanitaire) diffusers with Lamson blowers. R. Watson indicated that the trickling filter effluent pumps are a bottleneck at the plant, with a rating of 40 USmgd.

- The plant occasionally has experienced loses of the solids contactor MLSS, either due to shock loadings or a lack of flow. The lack of flow possibility may be comparable to low organic loading conditions at the West Windsor pilot unit.

- Air flow during our tour was set at 1228, 880 and 610 cfm for the re-aeration tank and the two contact tanks. The DO concentration based on the plants online equipment was 4 mg/L in the re-aeration pass and 2 mg/L in the second compartment.

- The plant recirculation pumps recirculate trickling filter effluent back to the tower at a rate of 4 USmgd.

- The aeration tanks are sloped to assist in removing snail shells out the aeration. Provision should be made to remove any shells that may accumulate in the aeration tank. Two to six foot snail deposits have been found in the aeration tank.

- The plant staff clean debris from the surface of the trickling filters once a month at the time when they also grease the bearings of the feeder mechanism.

- R. Watson outlined the means by which the plant controls the plant wasting. The system uses a Lotus 123 Macro to calculate the current day and past day's operating parameters and base the next day's wasting on an SRT of 3.5 d and a desired MLSS of 1200 mg/L. A seven-day average is used. The rate is constrained within a predetermined amount, and must remain within the limits of 35 to 165 USgpd. R. Watson demonstrated the Macros used by the plant, a copy of the outputs is attached. R. Watson was willing to supply the code to the City of Windsor if the plant proceeds with the TF/SC process.

- The original plant was a biological (6 USmgd rock TF) and physical-chemical facility with chemical precipitation, coagulation and flocculation (polymer addition), clarification, followed by carbon absorbers. Carbon absorbers were abandoned and the plant was upgraded to the TF/SC process.

- The plant uses old digesters for sludge holding tanks. Lime or ferric are used for sludge dewatering using plate and frame presses. The cake is hauled to landfill. The plant produces approximately 139 t/d of cake. The landfill site is 15 miles from the plant.

- The plant has 45 workers (20 operators, 21 mechanical, 2 dewatering operators and 2 administration). The onsite laboratory is used by both plants, the water department, and for industrial sampling and has 8 personnel. The onsite laboratory processes 4000 samples per month.

- The Duck and Rowlett Creek Plants are the only two plants in the area which pass the toxicity tests on the plant effluent. Daphnia and fat-head minnow tests are used. The plant dechlorinates the disinfected effluent.

- Plant sampling is based on composite samples for raw sewage and final effluent. Primary effluent is a grab sample.

- The plant has piston raw sludge pumps, R. Watson would prefer variable speed Moyno pumps with a Muffin Monster set-up. The biological sludge is fed to the primary clarifiers to co-settle and thicken with raw sludge and Rowlett Creek sludge. TF/SC biological sludge is continuously wasted using a controller on a number of control valves. The rate is between 55 and 165 USgpm.

- The CH2M HILL TUMMS Utility Management system is used for plant maintenance.

2) THE ROWLETT CREEK WPCP, CITY OF GARLAND, TEXAS

Date:	3 April 1991; 9:00 to 12:00
Contacts:	Mr. Lynn Norton, Director of Wastewater Mr. Rick Watson, Lead Process Manager Mr. Bill Pippen, Operations Supervisor
Attachments:	Rowlett Creek WPCP flow schematics and plant layouts; startup data and operational data from 1985 to 1991; plant 1990 Annual Report. These summaries are exceptional and should be reviewed with this narrative.
Plant Design:	16 USmgd; side-line equalization under construction; nitrification required.

Mr. Ron Seiger of CH2M HILL's Dallas office made local arrangements and accompanied our group on the tour. Mr. Rick Watson and Mr. Bill Pippin made introductory remarks to the group and lead us through the plant inspection. The plant superintendent is Mr. David Robertson. The following main items and discussion points were made:

- Hydraulic design capacity is 16 USmgd; effluent permit is 10 mg/L BOD_5, 15 mg/L SS plus nitrification to 2/5 mg/L NH_3-N in summer/winter. Summer is May to October. The plant achieves excellent nitrification.

- TF tower recirculation is provided by pumping to a maximum of 6 USmgd.
- Expansion is ongoing at the plant, including:
  - expanded flow equalization (side-line application) with 5 million gallon US storage after primary clarification to reduce diurnal peaks
  - one additional plastic media trickling filter and raising the media level on two existing TF's from 12 to 16 ft.
  - new fine bubble Sanitaire diffusers and Lamson blowers
  - secondary clarifiers with outboard weirs
- R. Watson clarified a question that was raised during the tour of the Duck Creek Plant concerning the difference between the RAS concentration and the concentration of the solids in the re-aeration tank. There is some mixing between the re-aeration tank and the second pass aeration tank, causing a reduction of the expected solids concentration in the re-aeration zone.
- G. Daigger noted that the Garland plants are very comparable to what is being tested at Windsor in terms of both organic and nitrogen removal in a single-stage trickling filter.
- There are four influents to the plant from the neighbouring areas. The influent flow rate is determined by a Parshall flume, the effluent by a Cipoletti weir at the chlorine contact chamber. Effluent re-aeration is practised by step aeration. Pista grit separators are used along with 3 covered primary clarifiers. Waste biological sludge is co-settled in the primaries.
- Due to an initial delay in the completion of the solids contact tanks, the plant began the TF/SC process by aerating the centre flocculating wells in the three final clarifiers (two used presently). The plant has maintained this aeration and believe it results in a further 2 ppm removal of soluble BOD_5.
- Snails accumulate in the plant, including SC aeration tanks and effluent polishing filters. Snail control and removal must be addressed in any design.
- RAS re-aeration is practised.

- The plant occasionally chlorinates the RAS to avoid filamentous growth and settling problems using about 100 lb Cl_2/1000 lb MLSS. A 'sporadic' F/M was also thought to be a source of settling problems. Staff believe that once the plant has additional equalization capacity that this problem may be eliminated, since it is not a problem at the Duck Creek Plant. The problem occurs once or twice a year especially in June or July. When SVI exceeds ~110, the plant examines sludge with a microscope and chlorinates if necessary.

- The plant must chlorinate year-round (State Health requirement). Although there is no requirement, the plant dechlorinates. CCC retention is 20 minutes at peak 2-hour flow.

- Snails are a problem at both Garland plants. At Duck Creek the level of snails is six feet in places. G. Daigger indicated that snails are a problem at the Salt Lake City plants as well. R. Watson recommended the ability to drain and remove snails as opposed to chlorination control. R. Watson was asked if flies were a problem around the trickling filters; he indicated that they are a problem at Rowlett Creek. This may be due to the lower wetting rate and the inability to recirculate.

- There have been diffuser fouling problems at the plant after five years, mainly due to underside fouling as a result of inadequate filtering of the air supply.

- The trickling filter towers have an internal temperature probe to monitor the temperature of this process.

- The Rowlett Creek WPCP intends to pump effluent approximately 15 miles to avoid a phosphorous removal requirement. Phosphorous removal would be required if the plant continues to discharge directly into the water supply reservoir.

- Historically the plant provided treatment in a five-foot trickling filter and solids stabilization.

- WAS returns upstream of the primary clarifiers. The plant aerates the trickling filter effluent channel to mix the trickling filter effluent with the RAS. Currently, two of the three final clarifiers are in operation.

- Rowlett Creek, like Duck Creek, has automatically backwashed sand filters on the secondary effluent. The plant shock loads the filters daily with chlorine to minimize or remove algal growth.

- Anaerobic digestion of raw and biological sludge is practised in 4 digesters. Only one digester is used presently; it is said to be operating at a 7-day HRT and giving a 60^+ percent VS destruction. A 27 000 PE chicken processing plant wastewater is received increasing raw influent volatiles concentration.
- The plant dewaters sludge with belt filter presses. Digester gas is stored in a sphere at 50 psi for flare-off or use in the plant's co-generation system. The system currently produces 40 percent of the plant's electrical energy requirements. Iron filters are used for H_2S removal along with numerous drip traps prior to a Cooper engine generator.
- Warm water fish have been observed in the plant outfall channel somewhat downstream of the plant.
- The plant also has a Dissolved Air Flotation unit, but it is not used. The plant has not found the DAF to increase the solids concentration and the unit is a large power user.

3) CENTRAL VALLEY WATER RECLAMATION FACILITY SALT LAKE CITY, UTAH

Date: 4 April 1991; 8:30 to 13:00 pm

Contacts: Mr. Reed Fisher, Assistant General Manager and Director of Operations
Mr. Bob Robertson, Lead Process Manager

Attachments: plant layouts; 1990 operational data; plant 1990 Annual Report; capital cost data.

Address:
800 West Central Valley Road
Salt Lake City
Utah 84119-3379
801-973-9100 fax 973-9100 ext. 125

Plant Design: 62.5 USmgd secondary; 100 USmgd pretreatment and pumping; no specific nitrification design.

Mr. Cliff Forsgren from CH2M HILL's Salt Lake City office made local arrangements and accompanied our group. The City's Mr. Reed Fisher and Mr. Bob Robertson introduced us to the plant and provided an extensive plant tour. The following main items and discussion points were made:

- The plant does obtain some nitrification, although it is not one of the plant's effluent requirements. The influent ammonia concentration is about 22 mg/L, and the trickling filter or solids contact effluent is generally 4 to 5 mg/L. Actual data received after the inspection are attached and should be reviewed.

- The plant's permit requires a maximum of 20 mg/L COD and 25 mg/L TSS. The plant currently treats approximately 47.7 USmgd. The secondary plant is designed to treat 62.5 USmgd with a 2.0 PF. Actual PF is 1.5 with a maximum observed flow of 85 to 90 USmgd observed on some holiday weekends. The plant's service area includes a population of 460,000 people and a small industrial component. The wastewater temperature varies from 27 °C in the summer to a low of 12 to 14 °C in the winter. The plant has a 160 acre site of which 50 acres is currently being used. The plant started operations in 1988.

- Pretreatment and influent pumping have been constructed for 100 USmgd capacity. This should be considered when reviewing capital costs. Headworks have four Infilco screens and room for two more. Motors were a problem on the screens because of low position at high water.

- Grit is dewatered using Wemco hydrocyclones, classified (effectively done) and conveyed to dumpsters. The plant's four Worthington raw sewage pumps provide a 30 foot lift; each pump is rated at 50 mgd providing double redundency. The pumps are variable speed units (VFD) with 300 HP right angled drives. The plant receives a 1.5 peaking factor, but was designed for 2.0. The greatest flows occur on holidays, especially Thanksgiving. On these holidays the plant will treat 85 to 90 mgd. The plant has grit capacity to treat 100 mgd.

- All plant channels were aerated (300 HP total) and were a significant fraction of the aeration volume in this plant, particularly for the TF/SC processes. The plant aerates the primary effluent channel leading to the trickling filters.

- The trickling filter is loaded at 40 lb BOD_5/1000 ft^{3}d. The trickling filter influent propeller pumps are rated at 36 USmgd and 300 HP. Staff would have preferred mixed-flow pumps here. The plant provides 150 percent recirculation around the trickling filters to maintain wetting rate.

Foaming was a problem in the recirculation well and required a spray to control.

- The trickling filters are 178 feet in diameter and 14 feet deep. The arms rotate at 1 revolution per minute on three of the filters, the other filters are rotating at 1 revolution every 2 minutes. Reversible arms or a mechanical drive are required for rpm control. Media was becoming brittle at the exposed top and there was concern about foot traffic on top for maintenance and cleaning. The plant is looking at tower covers to solve winter temperature loss, summer drying, and odour control. Temperature problems would probably best be addressed by closing vents.

- Some of the older trickling filters (6 years) are showing significant concrete deterioration around the vent openings, the newer filters (two years) have not been affected yet. Vent openings were very large and thought to cause unexpected temperature loss during winter, affecting unintentional nitrification. Despite this, winter effluent temperatures (typ. 14 °C but ranging as low as 12 °C) were not that low compared to expected Windsor conditions (low of 8 to 10 °C). Therefore, we should question the design temperature for the TF/SC recognizing that it was not intentionally designed to nitrify. Temperature loss needs to be minimized since a TF is essentially a cooling tower system as well. Filter walls were lined concrete block construction.

- The solids contact tank uses Norton fine pore dome diffusers. The four-pass solids contact tank has one re-aeration pass. HRT is designed at 75 minutes; it is not clear whether this included the channels. Channels were said to be 10 ft depth with coarse bubble aeration. Snail shells have been a problem in the aeration tank; when the power is off, the shells accumulate under the diffusers and in dead zones. Up to 3 to 4 ft deposits have been found. The MLSS concentration is currently 2800 mg/L; the target is 1800 mg/L. The RAS concentration is 8000 mg/L. The plant can get nitrification in the winter. The SC gives 4 to 5 mg/L NH_3-N reduction, but this decreases as loading increases.

- Secondary clarifiers are 125 ft diam. with 18-ft SWD. Flocculating centre well is 1/3 diam. and 11 ft deep. Plant has provision for alum and polymer addition in feed well. The feed well is basically a cup with an outlet at the top. Reed Fisher indicated that the secondary clarifiers can handle a maximum SOR of 600 USgpd/ft^2 and an average of 250 to 300 USgpd/ft^2. Each clarifier had an inlet weir control section. Plant information obtained on the site visit suggests that today's average flow

SOR is about 486 USgpd/ft^2 and that the design flow loading would be about 637 USgpd/ft^2. At the present time indications are that the plant may only add two more secondary clarifiers when secondary capacity is expanded to 100 USmgd average and 200 USmgd peak. This would result in peak SOR of 1630 USgpd/ft^2 (61 m^3/m^2-day or twice MOE design guideline for a nitrifying ASP). This situation should continue to be followed carefully prior to Windsor final design. Reed Fisher indicated that significant blanket level (10 to 12 ft) was carried in the clarifiers, but a sludge judge on our visit showed little blanket (6 in). Sludge is reported to settle well. Each clarifier had a separate RAS pump and underflow magmeter to a common return header to the SC. VFD driven pumps (20 HP) were set and operated at a constant speed. Design diurnal peak/average flow is 2/1 (actual is 1.5/1) as opposed to a wet weather peak flow situation in the dry Utah climate.

- Reed Fisher verbally indicated that plant cost was $130M ($98 M constr.) including land but spread over a 10-year period of construction. Therefore, costs would have to be compared on a present value basis. Life-cycle costs were said to be an important factor in choosing the TF/SC process as other facilities evaluated had lower initial capital cost. Cost data supplied at a later date is attached.

- Ferric is added to the plant's raw sewage, to protect the pumps from high levels of H_2S. The raw sewage also contains some inhibitors, although the concentration is acceptable.

- Waste biological sludge is thickened on three 2m Ashbrook-Simon-Hartley gravity belt thickeners. Wasting is continuous. Sludge is anaerobically digested. Four 625 kW cogen units heat and cool the entire plant.

- Plant dewaters to 10:00 p.m. Trucks to landfill are full in the morning. Plant has 5 trailers and 3 cabs. Landfill site is 7 miles. Belt presses deliver about 18 percent TS.

- Total plant staffing is 64 with 26 in operations and two shifts (5 in afternoon). Today O&M is said to be $6M annually. 75 percent of the plant's electrical energy is supplied through its co-generation system.

4) SALT LAKE CITY WATER RECLAMATION PLANT SALT LAKE CITY, UTAH

Date: 4 April 1991, 2:00 to 5:00 p.m.

Contact: Mr. William Farmer, Facilities Manager

Attachments: plant schematic and brief description

Address: 1365 West 2300 North
Salt Lake City, Utah 84116
801-799-4001

Plant Design: 56 USmgd; no nitrification mandate.

We had a thorough inspection of the SLC Water Reclamation Plant on the afternoon of 4 April 1991 with the assistance of the Facilities Manager Mr. William (Bill) Farmer. Mr. Cliff Forsgren of CH2M HILL's SLC office arranged for the inspection locally and accompanied us on the visit. This is a retrofitted TF/SC plant with a design hydraulic capacity of 56 USmgd. The following points summarize the inspection tour findings.

- Plant flow currently averages 35 USmgd with about seven percent industrial input. The plant maintains a 30 USmgd standby portable pumping capability. Permit limits are 25/25 monthly and 35/35 weekly maximums for BOD_5 and SS. A plant pamphlet indicated that the plant effluent quality ranges from 15 to 20 mg/L for BOD_5 and SS in winter to 10 to 15 mg/L in summer at the current plant flows (63 percent of design). This reportedly is achieved consistently with 15 percent of the treatment process units out-of-service.

- Plant primary effluent quality averages 110 mg/L BOD_5 (relatively high soluble fraction) and 60 to 70 mg/L SS, including internal plant recycle streams.

- The plant has eight 6-ft depth rock media trickling filters followed by a solids contact process. Three units have been taken off-line in the past to stress the remaining five units to determine the maximum sustainable organic load. The results showed that the plant could take 35 lb $BOD_5/1000\ ft^3.d$ during the tests up to the month of January. The conclusion was that the maximum sustainable organic load was 30 lb/1000 $ft^3.d$.

- The secondary clarifiers are 8 ft and 12 ft depths. The maximum recommended SOR is 600 USgpd/ft^2. A key design point was to provide a baffle skirt on the clarifier flocculating well and effective mixing of the RAS with the TF effluent. Plant peak flow is 40 USmgd with an average of 35 USmgd and a minimum of 20 USmgd.

- The trickling filters run with a controlled recycle to maintain the wetting rate. The plant staff feel it is important to maintain high hydraulic loading to control snails. Recirculation is controlled automatically. The plant has not experienced severe filter fly or odour problems to date.

- Discharge is to a 'Class 6' canal and then to Great Salt Lake. The plant will not likely ever have an effluent ammonia criteria, but it gets some nitrification in summer. This nitrification is not achieved at the simulated high loadings (30 lb BOD_5/1000 ft^3·d). If organic loading was reduced to the near 20 lb/1000 ft3.d, nitrification might be achieved. Minimum wastewater temperature is 8 °C after treatment. Ice buildup has been experienced on the TF distribution arm and ring, cables and turnbuckles. A crew is used to chip off these ice build-ups. A reduction in snail buildup has been experienced since startup of the SC unit; this is thought to be due to greater flushing action or to unobserved buildup in the SC and re-aeration units.

- The solids contact process use coarse bubble aeration in a long narrow channel with a HRT of about 20 minutes and 800-900 mg/L MLSS. During our inspection the air was reported to be at the minimum to avoid floc disruption. One observation was that the SC MLSS was noticeably dark in appearance. The SC process include return sludge re-aeration in 2 parallel tanks with a RAS concentration of 1800-2000 mg/L. Aeration was by WYSS diffusers. Aeration power for the SC and re-aeration channel was 100 HP. Design HRT in the re-aeration tank is 45 minutes. Screw pumps were used for RAS return to the SC contact channel. Design HRT in the SC unit is 15 minutes at design flow. Treatability results had indicated that a trickling filter alone would provide 25 mg/L SS with straight settling versus 15 mg/L SS with the solids contact process.

- Plant has two new final clarifiers at 65 ft diameter. Plant staff felt that clarifiers should have had surface skimmers.

- Plant power usage is 1200 kW; 900 kW for treatment; 300 kW for influent pumping

- Plant staff is 58 total with 20 maintenance, 25 operations, 8 lab, 5 office. Seventy-four industries are permitted to discharge to the plant; these are monitored by plant staff. Grounds maintenance is by contract.

- The City does toxicity testing of the treated effluent in-house. Main problem is with pH change during the testing. In September, toxicity limits will be placed on the plant; now just monitoring. Limit will be LC50

- Gravity thickening of primary raw plus waste biological is provided to yield about 6.5 percent sludge which is the maximum pumping capability at the plant. Thickener odour control is being evaluated to meet public perceptions. The unit runs at about 400 USgpd/ft^2 overflow rate. Sludge is about 2/3 raw and 1/3 biological. Biological is about 400,000 USgpd. Digested sludge is dewatered to 50 to 70 percent TS. Disposal costs are $35 US per ton to agricultural land, excluding capital. Raw sludge is screened using a 'HYCOR' automatic screen (J. Meunier); it is fairly dilute input then thickened in the gravity thickener. It is felt to improve digester performance.

- In 1990, the City embarked on a program to reuse treated effluent for a an enhanced wetland adjacent to the plant, possible freeway irrigation and gravel washing. Today, about 1 to 2 USmgd is diverted to the partially completed wetland as a trade-off for additional sludge drying bed area. We inspected the inlet area to the wetland only. It was apparent that some This suspended matter in the effluent was deposited in this area, suggesting that high quality effluents would be required for practical usage of this concept.

- Digester gas is used to power ~600 kW saving about $120,000/year electrical purchase. Engine heat and exhaust heat recovery is used for digester heating. Straight payback was said not to be good, but independent power was needed for backup supply. An NOX emission standard was set in SLC which had to be met (standard unknown). It was reported that Caterpillar engines had a problem meeting the standard or handling greater than about 1000 ppm H2S; Waukesha engines may be able to do better in connection with emissions. Farmer said that engine integrity would probably require < 500 ppm H2S anyway.

5) BOULDER 75th STREET WASTEWATER TREATMENT PLANT CITY OF BOULDER, COLORADO

Date: 5 April 1991

Contact: Mr. Floyd Bebler, Plant Superintendent

Attachment: 1989 Operating summary including plant flow schematic and monthly operational analysis. This attachment should be reviewed for a complete description of the plant design data.

Address:
Dept. of Public Works
Wastewater Treatment Plant
P.O. Box 791, Boulder, Co. 80306
Phone 441-3259

Plant Design: 16 USmgd; NH_3-N criteria will be reduced to 3 mg/L in near future; used for downstream irrigation.

Staff from CH2M HILL's Denver office arranged the site visit and accompanied us to the plant site.

- Process Flow: Parshall flume; Infilco climber screen (3/4"); aerated grit; primary sedimentation (4); intermediate P.S. to rock trickling filters c/w covers for odour control (scrubbers not used); interstage P.S. to 4-pass SC tank (3 passes used now; 2 in summer to try to avoid nitrification at this process stage); intermediate P.S. with ~40 percent of flow to separate stage plastic media nitrifying trickling filter (NTF); nitrified and non-nitrified effluents blended; disinfection with chlorine and dechlorination; effluent to Boulder Creek. Plant has 'Panalarm' display for indication but not control.

- This is a separate stage nitrifying trickling filter plant with discharge to the creek without post-clarification. The NTF has not yet developed a good biomass (started in May 1989). Plant gets ~15 BOD_5 and 6 to 8 SS. The NTF gives a range from 20 to 90 percent NH_3-N removal. NTF inlet BOD_5 and SS must be low to sustain separate stage nitrification. At the time of our visit the mixed effluent NH_3-N was ~8 mg/L from a raw influent of ~15 mg/L. Effluent temperature drops to ~9 °C in winter; nitrification is sustained. Some periods of weather below 0 °F are experienced, but 20 °F is normal low air temperature. The NTF tower had sample ports at 4 ft intervals, but the plant has little data. The top 8 ft were felt to be the most critical. Actual plant flows have averaged near 17 USmgd in summer months; the plant continued to meet effluent targets.

- Plant has operated SC unit with 500 to 2000 MLSS; typical is 1000 to 1400 mg/L; this must be controlled to avoid "bulking". SRT is targeted to 1 day to avoid nitrification in the SC and denitrification in the clarifier. There are 4 intermediate pumps feeding TF effluent to the SC unit (2 duty/2 standby), each with Louis Allis VFD drives. Soil conditions mandated extra pumping. SC (17 ft D) can be operated in parallel or plug flow; RAS re-aeration is not practised, but flexibility is available. Snails were said to be a "massive" problem. SC air diffusers are 18" above the floor and snails have been found up to that level. Tank dewatering capability is essential to handle snail problems, including good slopes and high volume flushing water with return to grit facility. Aeration was fine bubble tubular (rubber) with for 150 HP Lamson blowers and wall sound proofing. GE Fanuc single loop controllers on air lines. All channels (11 ft D) were aerated and formed a significant fraction of the effective aeration volume. Channel diffusers were said to be difficult to remove for maintenance.

- Three secondary clarifiers (National) have 16 to 17 ft SWD with vacuum draw-off (carry 1 ft blanket), flocculating centre well, four bridges, skimmers to inner weir on outboard system. Effluent was a brownish colour. Plant has experienced 'bulking' with effluent BOD_5 to 36 mg/L and SS to 20 mg/L during the event. Plant has operated with only one clarifier in service (all in-service during our visit). RAS with five 25/11.1 HP pumps; WAS with two 5 HP pumps.Fairbanks-Morse vertical turbine pumps lifted secondary effluent to NTF. Mist and foaming problems required control at this P.S.

- The NTF is flooded every 3 weeks to flush solids. Plant can recirculate around NTF. Additional NTF tower(s) will be required to meet more stringent NH_3 requirement.

- Nitrification was to prevent un-ionized NH_3 problems in the creek exacerbated by high stream pH (9.5) due to algal photosynthesis. Stream revegetation is being attempted to minimize high pH problems (shading). Plant effluent pH is ~ 7.2 typically.

- Two gravity thickeners are used to take primary and waste secondary sludge (sludge from rock trickling filters and solids contact process) to a concentration of 4 to 5 percent. These are followed by primary and secondary anaerobic digestion. Primary digester gives 20 to 25 day

Settleability of Solids, Item 19

Based on the results of the solids treatability task, the unit which had the best settling sludge based on the bench-top settling tests will rank a 10.

Dewaterability of Solids, Item 20

Based on the results of the solids treatability task, the unit which had the best dewatering sludge based on the bench-top dewatering tests will rank a 10.

OPERATIONAL CONCERNS (CATEGORY D)

Items 21 - 36

The fourth heading or category called operational concerns will include a number of items relating to the operation of each of the processes. These items will be ranked based on the pooled results of a questionnaire or ranking sheet completed by members of Steering Committee. Table 2 is a questionnaire for the items in this category, please complete the sheet and fax it to CH2M HILL by August 30, 1991 (fax no. (519)579-8986).. To assist in ranking the proposed plant expansion layout, the preliminary layouts for each process are attached (Figures 1 through 4). The list of items proposed for this heading include:

- operational ease
- operational flexibility
- environmental concerns
- operational knowledge of system in Ontario or City of Windsor
- expected land requirements or "foot-print" of proposed plant
- effluent aesthetics
- ability of process to be constructed in stages
- startup requirements of the process
- ability to control the process
- automation potential of the process
- ability of process to handle seasonal loadings
- operational sampling requirements
- potential for operational problems
- potential for maintenance problems
- odour potential
- noise potential

The weighting factor on each of these items will vary depending on the items's expected importance; however, these items are expected to have a weighting factor of between one and two (based on a weighting of 1, total of 160).

ECONOMICS (CATEGORY E)

Items 37 - 39

The fifth heading entitled economics will rank the units based on the initial capital cost, yearly operational and maintenance costs, and the amortized value of the project over 20 years. This heading will not be completed until the costing has been completed by LaFontaine, Cowie, Buratto and Associates Ltd. The cost of the full-scale plant will be critical to the City; therefore a weighting factor of ten is proposed (for a total of 300).

The overall ranking will have a possible total of 1,000 points, with 360, 90, 90, 160, and 300 points proposed for each major category (see Table 1).

TABLE 1 – RANKING OF PILOT PLANT PERFORMANCE

PROCESS	CAT 'A' EFFLUENT PERFORMANCE														CAT 'B' HYD. PERF.			CAT 'C' SOLIDS PROD			CAT 'D' OPERATIONAL CONCERNS																CAT 'E' ECONOMICS			TOTALS					
	MONTHLY COMPLIANCE					PROB. LESS THAN OBJECTIVE					PROB LESS THAN UPPER																																		
	TSS	BOD5	TP	NH_3-N	UN-IONIZED NH_3-N	TSS	BOD5	TP	NH_3-N	UN-IONIZED NH_3-N	TSS	BOD5	TP	EFF. TOXICITY	AVG Q	DURNIAL Q	PEAK DAY Q	SOLIDS YIELD	SETTLER CHAR	DEWATER CHAR	OPER. EASE	OPER. FLEX	ENVIR. CONCERNS	SYSTEMS KNOW	PLAET SIZE	EFFUENT AESTHETIC	CONST IN STAGES	STARTUP EASE	CONTROL	AUTOMATION	SEASONAL LOADINGS	SAMPLING REQ'TS	OPER. PROBLEMS	MAINT. PROBLEMS	ODOURS	NOISE	CAPITAL COST	O & M COST	AMORTIZED COST	CATEGORY A	CATEGORY B	CATEGORY C	CATEGORY D	CATEGORY E	TOTAL
WEIGHTING 100 %	5 50	5 50	5 50	5 50	5 50	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	3 30	3 30	3 30	3 30	3 30	3 30	3 30	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	10 100	10 100	10 100	360	90	90	160	300	1000
TF/SC																																								0 0	0 0	0 0	0 0	0 0	0
RBC																																								0 0	0 0	0 0	0 0	0 0	0
BAF																																								0 0	0 0	0 0	0 0	0 0	0
ASP																																								0 0	0 0	0 0	0 0	0 0	0

Table 2
Questionnaire for Operational Concerns

Item	Processes *				Weighting
	TF/SC	RBC	BAF	ASP	
Operational ease					1
Operational flexibility					1
Environmental concerns					1
Operational Knowledge of System in Ontario or City of Windsor					1
Expected Land Requirements or "foot-print" of proposed plant					1
Effluent aesthetics					1
Ability of process to be constructed in stages					1
Startup requirements of the process					1
Ability to control the process					1
Automation potential of the process					1
Ability of process to handle seasonal loadings					1
Operational sampling requirements					1
Potential for operational problems					1
Potential for maintenance problems					1
Odour potential					1
Noise potential					1
Note: To be completed and faxed to CH2M HILL by August 30, 1991					

* Each process to be ranked from 0 to 10, with 10 being the best.

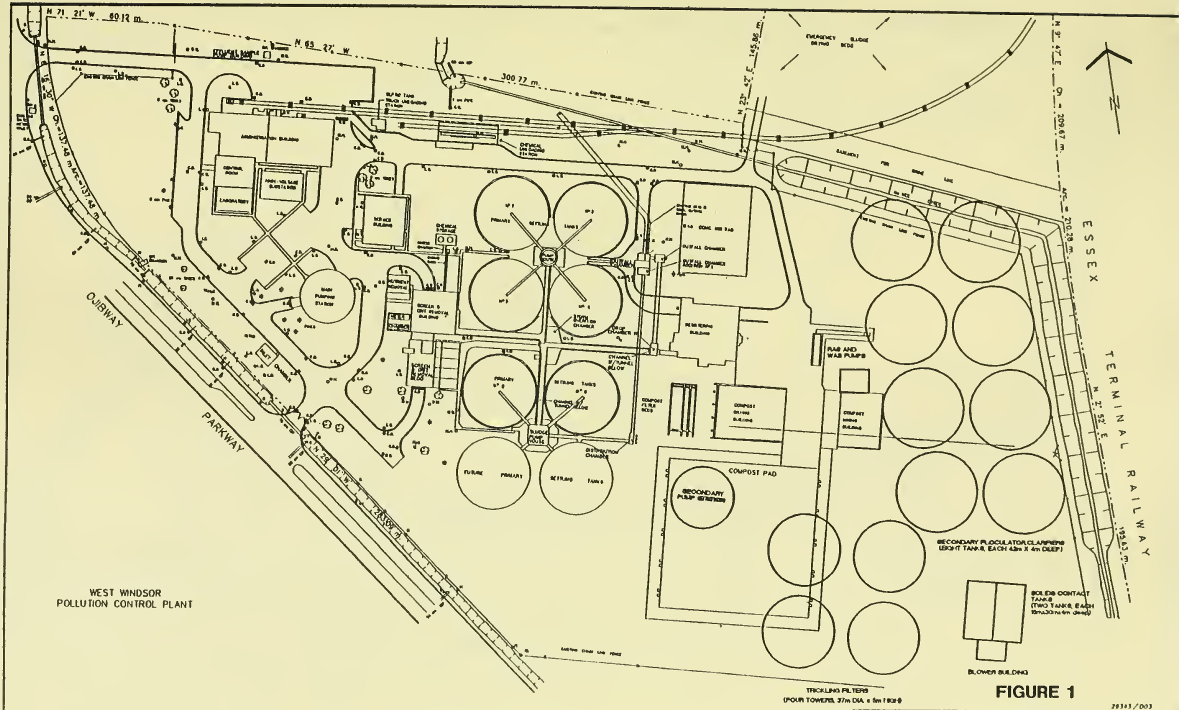

FIGURE 1

29343/003

WEST WINDSOR
WATER POLLUTION CONTROL PLANT

POSSIBLE LAYOUT
FOR THE WEST WINDSOR PCP UPGRADE PROCESS
TRICKLING FILTER/SOLID CONTACT

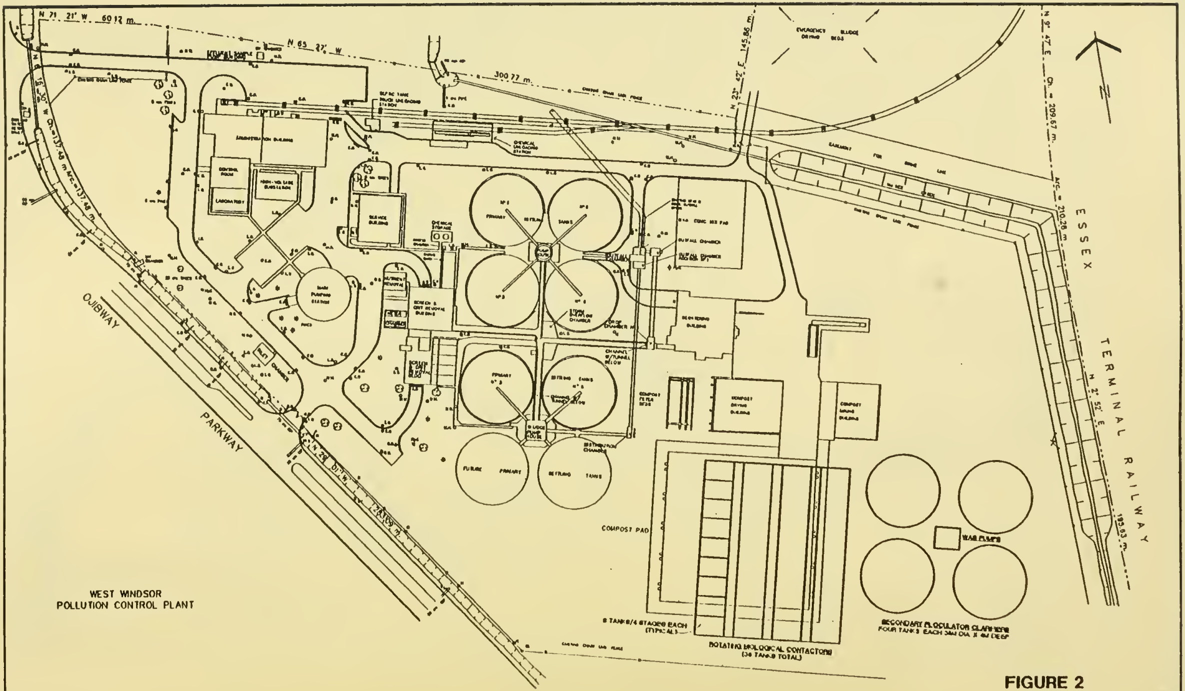

FIGURE 2

29143 / 29311004

WEST WINDSOR
WATER POLLUTION CONTROL PLANT

POSSIBLE LAYOUT
FOR THE WEST WINDSOR PCP UPGRADE PROCESS
ROTATING BIOLOGICAL CONTACTOR

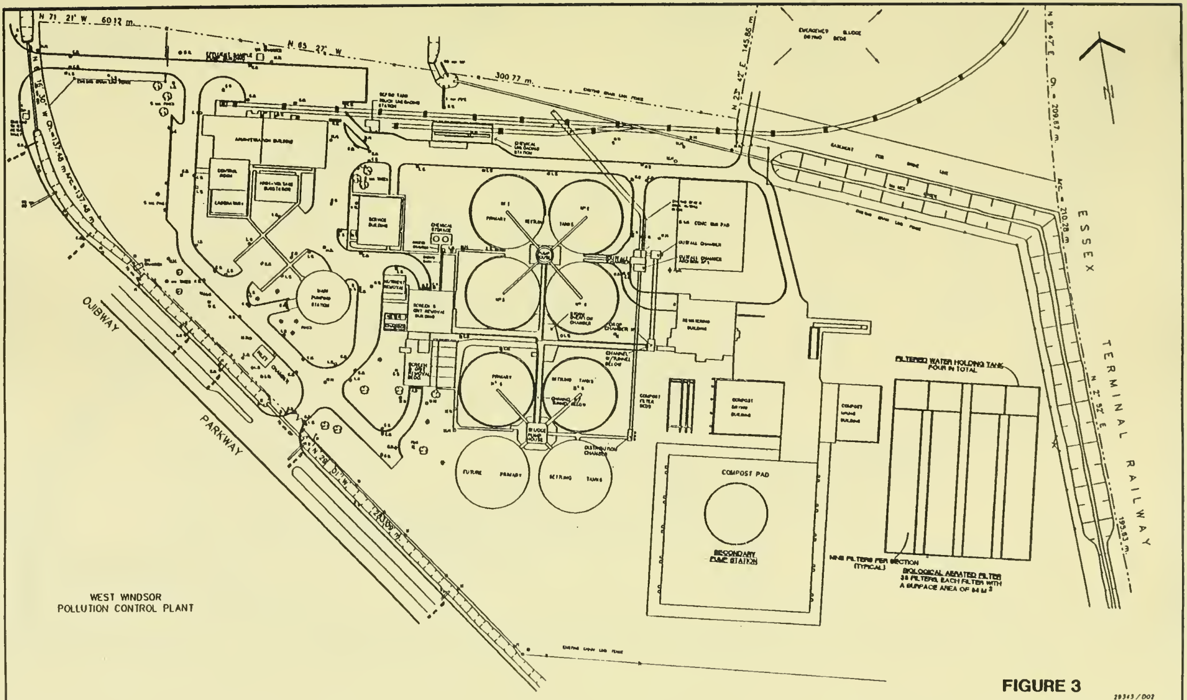

FIGURE 3

29343/002

WEST WINDSOR
WATER POLLUTION CONTROL PLANT

POSSIBLE LAYOUT
FOR THE WEST WINDSOR PCP UPGRADE PROCESS
BIOLOGICAL AERATED FILTER

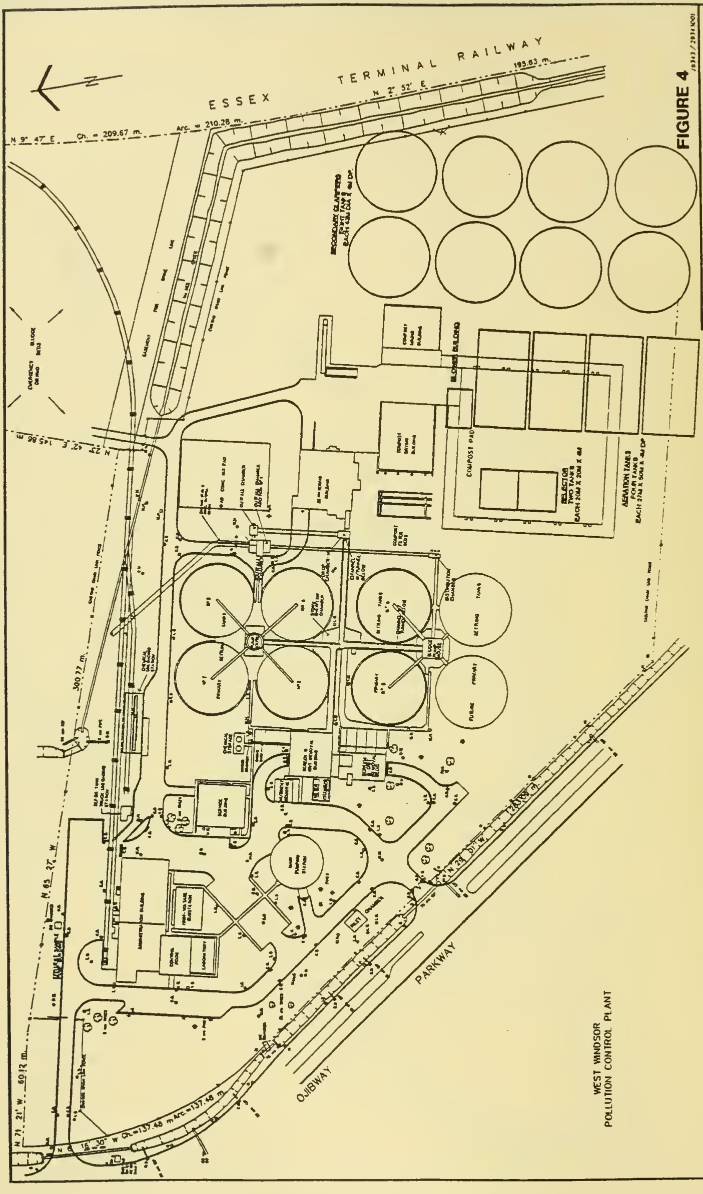
ESSEX TERMINAL RAILWAY
N 2° 52' E
193.63 m.
Arc. = 210.28 m.
N 9° 47' E
Ch. = 209.67 m.
SECONDARY CLARIFIERS EIGHT TANKS EACH 43M DIA X 4M DP.
EMERGENCY SLUDGE DRYING BEDS
BLOWER BUILDING
COMPOST PAD
SELECTOR TWO TANKS EACH 20M X 20M X 4M
AERATION TANKS FOUR TANKS EACH 27M X 50M X 4M DP
N 22° 42' E 145.86 m
300.77 m.
N 65° 27' W
N 71° 21' W 60.12 m.
N 6° 15' 30" W Ch.=137.48 m Arc.=137.48 m
N 29° 01' W
PARKWAY
OJIBWAY
WEST WINDSOR POLLUTION CONTROL PLANT

FIGURE 4

(8343/2834300)

WEST WINDSOR
WATER POLLUTION CONTROL PLANT

POSSIBLE LAYOUT
WEST WINDSOR PCP UPGRADE PROCESS
'NON-TRADITIONAL' ACTIVATED SLUDGE

Appendix J

COMPARATIVE COST ESTIMATES

WEST WINDSOR POLLUTION CONTROL PLANT

INNOVATIVE TREATMENT ALTERNATIVES

FULL-SCALE PROCESS COST ESTIMATES

WEST WINDSOR POLLUTION CONTROL PLANT

FULL SCALE PROCESS COST ESTIMATES

BACKGROUND

This appendix to the report on "Evaluation of Innovative Alternatives for the West Windsor PCP Upgrade" contains only cost comparison information. Comments and recommendations on the comparitive merits of the alternative processes are contained in the Executive Summary Conclusions and Recommendations and in the body of the report.

The West Windsor Pollution Control Plant is located on a 14.6 hectare (36 acres) site at the corner of Ojibway Parkway and Sandwich Street. The original plant went into operation in 1970 as a primary treatment facility with the rated capacity of 109,000 m^3/d (24 MIGD). The level of treatment was upgraded to "physical - chemical" in 1973 to meet Provincial phosphorous removal requirements. The plant was expanded in 1980 to the present capacity of 159,000 m^3/d (35 MIGD).

The original plant layout was based on the ultimate provision of a 273,000 m^3/d (60 MIGD) conventional activated sludge treatment plant on the 14.6 hectare site. Existing raw sewage pumping, screening and grit removal facilities are sized for the ultimate plant capacity. Final settling tank sizing at the time of the original design was based on providing the same surface area as provided for the primary settling tanks. Using current design standards for activated sludge plants, the surface area of the final settling tanks will be approximately twice the surface area of the primary tanks. As a consequence, there is not sufficient space on the original site to accommodate the ultimate plant capacity and it will be necessary to locate some tankage on City owned property on the south side of Ojibway Parkway. For the purposes of this study, it is assumed final settling tanks will be located on the south side of the Parkway and biological treatment tankage on the original site.

FULL SCALE PLANT LAYOUTS

The following full scale treatment alternatives have been considered in this study.

- Conventional activated sludge
- Modified activated sludge

- Trickling filter/solids contact (TF/SC)
- Rotating biological contactor (RBC)
- Biological aerated filter (BAF)

Tankage sizing requirements for each of these alternatives are summarized in Tables 1 - 5 respectively.

Plant layouts for each of the alternatives are shown in Figures 1 - 5 respectively.

The following points should be noted with respect to tankage sizing information.

- All tankage sizes and configurations were selected for an ultimate plant capacity of 273,000 m^3/d. In some cases this results in tankage modules that are slightly oversized or undersized for the 159,000 m^3/d plant. The discrepancies in sizing are marginal and do not significantly effect cost estimates.
- The conventional activated sludge alternative is based on MOE Design Guidelines calling for 6 hours hydraulic retention time (HRT) in the aeration tanks at design dry weather flow (DWF).
- The modified activated sludge alternative has a 3 hour HRT in the aeration tanks at design DWF.
- Final settling tanks for both the conventional and modified activated sludge processes are in accordance with MOE Design Guidelines for plants with a nitrification requirement.
- Both activated sludge treatment alternatives include anoxic selector tanks. Approximately 3,000 m^3 of selector volume is included for the 159,000 m^3/d plant and 5,300 m^3 for the 273,000 m^3/d plant.
- The RBC treatment process description shows two alternatives for sizing of final settling tanks. One alternative is based on conventional MOE Design Guidelines and results in 13 - 45 m diameter tanks being required for a 273,000 m^3/d plant. By comparison, tankage sizing based on criteria from the innovative test program results in 6 - 45 m diameter tanks.

  The MOE has reviewed the test program findings and has indicated acceptance of a peak clarifier surface overflow rate of three times the MOE Design Guidelines would be contingent on a process audit to establish the validity of the criteria for a full scale plant. The design layout, therefore, shows both clarifier design alternative and provides sufficient space for tankage to MOE Guidelines.

- The TF/FC alternative includes solids contact tankage sized for a 48 minute HRT at design DWF. In fact, it is likely a good portion of the solids contact volume could be provided in aerated channels conveying mixed liquor to the final settling tanks.

- One additional primary clarifier is required for the BAF alternative in the ultimate plant layout to account for the 13% backwash recycle stream which will be returned to co-settle with the raw sewage in the primary settling tanks. For the initial capacity of 159,000 m^3/d, the 13% recycle stream will result in a peak surface overflow rate of 62.65 m^3/m^2 day on the existing 6 primary settling tanks.

- The original conventional activated sludge plant layout included smaller final settling tanks and incorporated all tankage on the original site. An hydraulic analysis has confirmed that, with efficient hydraulic design, gravity flow would have been possible through the original plant layout. However, using large final settling tanks as required by today's design standards makes it necessary to construct part of the tankage on City owned property south of Ojibway Parkway. This separation of tankage and less compact layout eliminates the possibility of gravity flow and results in the need for additional pumping facilities.

All treatment alternatives, with the exception of the BAF process, require pumping facilities. For the purposes of this study, it is assumed the pumping facilities will be positioned to pump primary effluent for subsequent gravity flow through the biological portion of the treatment process.

During final design, detailed hydraulic calculations and complete information on soil conditions should be evaluated to select the best position for the pumping facilities both in terms of capital and operating costs. Primary effluent pumping would appear to offer a capital cost advantage by elevating downstream tankage to take advantage of foundation conditions which are best near the ground surface and deteriorate with depth. Final effluent pumping may achieve some operating cost savings if hydraulic calculations show gravity flow is possible during low flow/low river level conditions.

Primary effluent pumping will definitely be required for the TF/SC alternative.

CAPITAL COST ESTIMATES

Capital cost estimates have been developed using information from a variety of sources including:

- Actual costs for construction of activated sludge treatment facilities at the Little River Pollution Control Plant in the City of Windsor during the period 1987 - 1991.
- Literature information on capital costs and cost comparisons for activated sludge, TF/SC, RBC and BAF treatment facility. Literature sources include:
    - A U.S. Army Corps of Engineers August 1988 publication titled "The Trickling Filter/Solids Contact Process: Application to Army Wastewater Plants".
    - A 1981 Report by Brown and Caldwell titled, "Engineering - Economic Comparison of the TF/SC Process to Conventional Technology".
    - A 1992 Associated Engineering/Brown and Caldwell/Reid Crowther Report for the Greater Vancouver Regional District - Annacis Island and Lulu Island Wastewater Treatment Plants.
    - A December 1988 paper by Dillon and Thomas titled, "A Pilot-Scale Evaluation of the 'Biocarbon Process' for the Treatment of Settled Sewage and for Tertiary Nitrification of Secondary Effluent".
    - An April 1990 Report by Montgomery Consulting Engineers - "Technology Assessment of the BAF".
- Cost information for construction of BAF facilities in Sherbrooke, Quebec and TF/SC facilities in Central Valley, Utah.
- Cost information from RBC and BAF equipment suppliers.

Estimated capital cost comparisons for the five (5) treatment alternatives are presented in Table 6 for the 159,000 m^3/d plant and in Table 7 for the 273,000 m^3/d plant. **THE FOLLOWING CONDITIONS AND QUALIFICATIONS MUST BE TAKEN INTO CONSIDERATION WHEN REVIEWING THESE COST COMPARISONS:**

- Facilities common to all alternatives are not included in the cost estimates. Examples of common elements include; chlorination - dechlorination facilities;

site work including access roads, drainage, and process sewers; upgrading and expansion of the existing plant electrical distribution and instrumentation and control systems; potable and process water facilities.

- No allowance is included in the costs for known poor soil conditions at the West Windsor site. Cost estimates for the activated sludge alternatives are based on Little River Pollution Control Plant construction costs where the soil bearing capacity is 3,000 lbs./ft.^{2} or greater. Soil bearing capacities at the West Windsor site are more typically 2,000 lbs./ft.^{2} or less and vary with depth and location. Special provisions are necessary at the West Windsor site to accommodate differential settlement.

- No costs are included for standby electric power generator facilities.

- No allowances is made for any changes in sludge handling, digestion or dewatering.

- The BAF sizing information and layout shows spare cells to allow for cell cleaning. The spare cells are not included in the cost estimate since other alternatives do not include costs for backup capacity.

- No allowance is made for the possible need for grease/oil removal prior to the BAF units. Indications are that yearly cleaning of the cells will suffice.

- Primary effluent pumping facilities are assumed with a 15 foot static lift for the activated sludge and RBC alternatives and a 20 foot static lift for the TF/SC alternative.

- No costs are included for administration building or laboratory facilities.

- In general, the cost estimates assume open tankage without covers. For the BAF alternative the filter surface area will be open with pipe galleries and service areas enclosed. For the RBC process uninsulated fiberglass covers are included for the RBC units.

- Land costs are not included in the estimates. Land area requirements are generally the same for all alternatives with the exception of the BAF process. The land area south of Ojibway Parkway will not be required for treatment facilities if the BAF process is selected.

- The cost estimates do not include any allowance for engineering, contingencies, taxes or interim financing.

retention time. Two Humboldt centrifuges 'thicken' digested sludge to ~10 percent TS (higher solids too difficult to handle; some centrifuge plugging was experienced) which is still suitable for land utilization: 60 to 90 day sludge storage is provided. Sludge has been designated (certified) as an organic fertilizer. Digester supernatant and centrate are equalized prior to return to the rock trickling filters.

- Power outages are a problem. The plant load management system is not entirely reliable. A co-generation system is provided and is interconnected to the grid. Power is sold at ~4.25 cents US/kWh. Waste heat is used for digester and building heating.
- When an impending new effluent permit is delivered, effluent biomonitoring will be required. The plant has had few exceedances in this area; exceedances have generally been due to 'high' NH_3. The receiving stream is a warm water habitat.
- Plant operational staff had design input at the 40, 60 and 80 percent design stages.

6) LITTLETON/ENGLEWOOD REGIONAL WASTEWATER TREATMENT PLANT
ENGLEWOOD, COLORADO

Date: 5 April, 1991; 2:20 p.m.

Contact: Mr.Dennis Stowe

Attachment: description of original plant located at this site; plant is currently in re-construction with addition of new TF/SC process.

Plant Design: 37 USmgd; partial seasonal nitrification.

- Feed to the plant is from the two cities named above. A new 20 USmgd pump station was under construction; however the plant design capacity was 37 USmgd. The existing plant bypasses around the existing rock trickling filters on a regular basis.
- Relatively little information was available from this plant because of the reconstruction activity and time available with plant staff. This plant was in some difficulty operationally because of reconstruction activity. One new plastic media TF/SC unit is in construction in parallel with an existing air activated sludge facility. Ultimately, three additional TF/SC

units may be built. The TF tower is designed with complete backwashing capability for snail control. Backwash water will be chlorinated to kill snails and then pumped to digestion.

Effluent objectives will be for nitrification with 5.4 mg/L in summer and 13 mg/L winter NH_3-N limits. These values were developed based on a receiving stream study.

The existing activated sludge plant consisted of five parallel trains with GRAY/Sanitaire diffusers. MLSS was high at about 6000 mg/L. Secondary clarifiers showed evidence of a severe incidence of Nocardia type foam/scum with much floating scum on the surface of the clarifiers. An interesting problem observed at the plant was the RAS wet well; it was constructed for 8 clarifiers but only four clarifiers were installed. This resulted in dead zones in the over-sized wet well.

Please contact CH2M HILL for correction of errors or omissions in these notes.

Site Visit Summary prepared by:

CH2M HILL ENGINEERING LTD.

Joe P. Stephenson, P.Eng.

TENTATIVE SCHEDULE
revised April 1, 1991

WEST WINDSOR PCP UPGRADE

TF/SC SITE VISITS - APRIL 2 TO APRIL 6 ,1991

TENTATIVE SCHEDULE

DAY	EVENT	TIME	DESCRIPTION
TUES. APRIL 2	FLIGHT TO DALLAS	9:00 am DETROIT	DL 1271 FLIGHT DETROIT TO DALLAS DEPARTS 9:00 am ARRIVES 10:45 am
	DRIVE TO GARLAND	11:15 am	DRIVE FROM DALLAS TO GARLAND, 1 h 15 min TRIP (11:15 - 12:30)
	LUNCH	12:30 pm	QUICK LUNCH (12:30 - 1:00)
	PLANT VISIT	1:00 pm	VISIT DUCK CREEK GARLAND WPCP (1:30 - 4:30)
	DINNER	6:00 pm	
	HOTEL	8:00 pm	HOLIDAY INN NORTHEAST ON JUPITER RD, GARLAND, PHONE (214) 341-5400
WED. APRIL 3	BREAK	7:30 am	BREAKFAST FROM 7:30 TO 8:30 am
	PLANT VISIT	9:00 am	VISIT ROWLETT CREEK GARLAND WPCP (9:00 - 12:00)
	LUNCH	12:15 am	LUNCH FROM 12:15 - 1:15 am
	DRIVE TO DALLAS	1:15 pm	DRIVE TO DALLAS AIRPORT (1:15 - 2:30 pm)
	FLIGHT TO SALT LAKE CITY	3:45 pm	DL 1735 FLIGHT FROM DALLAS TO SALT LAKE CITY DEPARTS 3:45 pm, ARRIVES 5:17 pm
	DINNER	7:00 pm	
	HOTEL	9:00 pm	OLYMPUS HOTEL, 161 WEST 600 SOUTH, SALT LAKE CITY, PHONE (801) 521-7373

TENTATIVE SCHEDULE
revised April 1, 1991

DAY	EVENT	TIME	DESCRIPTION
THUR. APRIL 4	BREAK	7:30 am	BREAKFAST FROM 7:30 TO 8:30 am
	PLANT VISIT	9:00 am	VISIT CENTRAL VALLEY WPCP (9:00 - 11:45 pm)
	LUNCH	12:00 pm	LUNCH FROM 12:00 - 1:30 pm
	PLANT VISIT	1:45 pm	VISIT SALT LAKE CITY WPCP (1:45 - 4:30 pm)
	DINNER	6:00 pm	DINNER AT AIRPORT (6:00 - 7:15 pm)
	FLIGHT TO DENVER	7:54 pm	DL 1298 FLIGHT FROM SALT LAKE CITY TO DENVER, DEPART 7:54 pm, ARRIVES AT 9:20 pm
	HOTEL	10:00 pm	LANDMARK BEST-WESTERN HOTEL, 455 SOUTH COLORADO BLVD, DENVER, PHONE (303) 388-5561
FRI. APRIL 5	BREAK	7:00 am	BREAKFAST FROM 7:00 TO 7:45 am
	DRIVE TO BOULDER	8:00 am	DRIVE TO BOULDER WPCP
	PLANT VISIT	9:30 am	VISIT BOULDER WPCP (9:30 - 12:00)
	LUNCH AND DRIVE	12:15 pm	QUICK LUNCH (12:00 - 12:30 pm) AND DRIVE BACK TO DENVER (12:30 - 1:30 pm)
	PLANT VISIT	2:00 pm	VISIT LITTLETON-ENGLEWOOD WPCP (2:00 - 5:00 pm)
	DINNER	7:30 pm	DINNER
	HOTEL	9:00 pm	LANDMARK BEST-WESTERN HOTEL, 455 SOUTH COLORADO BLVD, DENVER, PHONE (303) 388-5561
SAT. APRIL 6	BREAK	7:00 am	BREAKFAST FROM 7:00 - 7:30 am
	FLIGHT TO DETROIT	9:10 am	DL 334 FLIGHT FROM DENVER TO DETROIT (WITH STOPOVER IN CINCINNATI) DEPARTS 9:10 am, ARRIVES CINCINNATI 1:40 pm, DL 3250 DEPARTS 2:20 pm ARRIVES DETROIT AT 3:40 pm

CITY OF WINDSOR EVALUATION OF
INNOVATIVE ALTERNATIVE FOR THE
WEST WINDSOR PCP UPGRADE

SITE VISIT

DATE: May 3, 1991

LOCATION:

PROJECT: ONT29343.C0

ATTENDEES:

Mr. L. Romano, City of Windsor
Mr. H. Horneck, LCBA
Mr. M. Newbigging, CH2M HILL
Mr. M. Maki, City of Windsor
Mr. J. Stephenson, CH2M HILL

SUBJECT: Summary of Site Visit to BAF Facilities in Quebec

A site visit to two BAF (Biocarbone) units located in Quebec was conducted on May 3, 1991. Both plants are designed for organic or BOD_5 removal only. The following memorandum summarizes the main points discussed during the trip.

Sherbrooke, Quebec

The Sherbrooke plant was re-visited since the plant is now starting up. The City and CH2M HILL visited the plant a year ago (May 8, 1990) when the plant was still under construction. Representatives from John Meunier Inc. (Mr. F. Seguin, Mr. H. Van Hoi and Mr. M. Emond) and Le Groupe Teknika (Ms. A. Kantardjieff) showed our group around the plant. The plant had just been started and was going through initial testing. The plant was mainly treating clean water. Our group was toured through the entire facility, the main points discussed were:

- Plant capacity is rated at an average daily flow of 80,000 m^3/d (17.6 MIGD) with a peak capacity of 125,000 m^3/d (27.5 MIGD). The raw wastewater is expected to have a total BOD_5 concentration of approximately 100 mg/L. The plant effluent criteria is 20, 20, and 0.8 for suspended solids, total BOD_5 and total phosphorus.
- Sludge from the plant will be lime stabilized. The plant has a slurry tank into which the plant can treat thickened or co-thickened sludge. Lime can also be used to adjust the pH of the raw sewage.

- Sixty percent of the flow to the plant is by gravity. In addition, two main pumping stations discharge to the plant. The influent works has stainless steel gates, with the ability to control bypass flows.

- The plant has J. Meunier raw screens with 0.5 inch openings. Screens can be controlled by headloss across the screen or on a time elapse basis. The number of screens in service is controlled by the flow rate. Belt conveyors are used for removal of the screened material.

- Aerated grit tanks are used for grit removal. Tanks also has surface skimmers for grease removal. Air flows are regulated along the tanks. High air flow are maintained at head end of the tank, tapering to low air flows at the effluent. Grit tanks are enclosed in a building, all equipment is explosion rated.

- The primary clarifiers are also covered, although not enclosed. The primary clarifiers are used to settled raw sewage and can be used as thickeners (obtained up to 7% TS). Clarifiers are "tube settlers", manufacturer is Degremont. Tanks are provided for chemical and polymer, with mixing tanks for addition prior to primaries. System appears to be very flexible for chemical and polymer addition. In terms of the operation of the thickeners, backwash water can be thickened then proceed to primaries or directly to presses.

- The BAF (Biocarbone) process is enclosed in a building and each train (there are two) of BAF cells is enclosed in a fibreglass encloser. Air is supplied by five blowers, three are centrifugal and two are positive displacement. The centrifugal blowers are Hoffman 300 hp units, with variable frequency controllers and auto suction throttling capability. The centrifugal blowers are used primarily for process air. The positive displacement blowers are used for backwashing. The blowers are provided with insulating curtains for noise reduction. The inlet gates to the BAF cells are controlled. Flow passes over a plate to dissipate the flow.

- During our walk by the BAF cells, the unit was being backwashed. Our group was able to observe the various cycles of air, water and air and water. The operation of the three syphons, located on one wall of the tank, was also observed by the group. It was noted that a significant amount of media was being backwashed out the tank and into the backwash channel. Meunier indicated that this occurs during the initially cleaning cycles, since some of the media fines float on the liquid surface. The large media in the channel is pumped back into the tank after backwashing.

- Our group inspected the backwash pumping gallery. The backwash pump was exhibiting a good deal of vibration, Mr. M. Emond indicated that this problem was being corrected.
- Thickened sludge from the plant is pumped through two plate and frame filter presses. Pumps for sludge feeding and chemical and polymer addition are located in an adjacent gallery.

Boisbriand, Quebec

Mr. F. Seguin, Mr. H. Van Hoi of J. Meunier also took our group on a brief tour of the Boisbriand plant located north of Montreal. The Boisbriand plant is rated at 25,000 m^3/d (5.5) MIGD. The plant is operated partially by J. Meunier, during the tour we met Mr. C. Meunier, who is in charge of the plant operations. The following items were discussed:

- The control room was visited. The computer system provides current status of the BAF unit and trending data. The plant was in the process of conducting an air scouring when we arrived. The air scouring was being used by the operators to breakup the media and move the solids further into the bed. The operators use this procedure to avoid unnecessary backwashing.
- The plant has tube settlers, similar to Sherbrooke. The clarifiers are Degremont equipment and were constructed from the existing "regular" primaries during the last expansion.
- The BAF (biocarbone) system used at this plant is different from Sherbrooke. The cells (eight) are pie shaped and form a circle. Influent flows in on the outside of the circle, effluent flows into the centre of the circle. The water backwashing is performed by a mechanism in the centre well which seals the effluent standpipe and pumps the effluent through the cell being backwashed. The suppliers indicated that they use this setup for smaller plant, and are currently planning a similar facility for Lavel, Quebec. The proposed Lavel plant will have two circular BAF units, each approximately a third larger than Boisbriand's cells.
- The plant was also using an influent online TOD meter, which J. Meunier distributes, to control the process air being used at the plant, in an attempt to minimize air usage and excessive backwashing.

Appendix I

COMPARISON PROCEDURE

MEMORANDUM

TO: City of Windsor

FROM: CH2M HILL ENGINEERING LTD.

DATE: August 20, 1991

SUBJECT: Comparison Ranking for Pilot Plant Alternatives for the West Windsor PCP - Revised

PROJECT: ONT29343.A0

To assist in determining the best alternative treatment process for the West Windsor PCP, the following comparison approach will be used. The comparative approach will involve ranking each of the four processes on a number of different items. The different items will be ranked based on the performance of each pilot plant and its applicability for use at the West Windsor PCP. The items can be grouped into the following main headings or categories:

- Effluent performance
- Hydraulic performance
- Solids handling and solids yield
- Operational concerns
- Economics

The first three headings will be determined based on the performance of the pilot plants over the eleven-month pilot plant operating period. The fourth heading, operational concerns, will be accessed based on the results of a questionnaire, to be filled out by various Steering Committee members. The questionnaire is attached and will be discussed under operational concerns. The last heading, economics, will be addressed by Lafontaine, Cowie, Buratto and Associates Ltd., with assistance from CH2M HILL.

Each item will be ranked from 0 to 10, with 10 being the highest score. Each item will then be weighted to indicate the importance of the item to the overall selection. The totals for each process will be used to determine the best alternate for the West Windsor PCP.

The following section will discuss the means being used to rank the pilot plants for the first three major headings. Each major heading is broken into various items and ranked separately. A blank spreadsheet indicating the items and the presently proposed weightings is shown in Table 1.

EFFLUENT PERFORMANCE (CATEGORY A)

Monthly Compliance, Items 1 - 5

The monthly averages for each process, for the period when the unit was operating as designed, will be calculated and ranked based on the number of months each unit met the monthly average compared to the total number of months operated. Therefore, a unit that met the average monthly effluent SS objective, 7 out of 10 months would score a 7. The monthly effluent objectives were 15, 15, 0.5, 0.1 and 2 mg/L for total BOD_5, SS, TP and un-ionized ammonia and ammonia-N. The weighting value of five is proposed for this item, since this is a major criteria of the proposed plant (total of 250).

Probability Less Than Objective, Items 6 - 10

These items were chosen to indicate which unit was meeting the effluent objectives on a daily basis. For these items the percent of the time the pilot produced an effluent less than the objectives listed above will be used to rank each unit. For example, if the pilot's effluent SS concentration is less 15 mg/L, 90 percent of the time, the process will rank a 9. It is proposed that a weighting for these items be set at one, since this was not an effluent criteria, but does indicate which unit will be most likely to meet the plant's objective with a degree of confidence (total of 50).

Probability Less Than Policy 08-01, Items 11 - 13

These items are similar to the items above. except they are compared to upper limits, in this case to Policy 08-01 limits. These limits are 25, 25 and 1 mg/L for total BOD_5, SS and TP. These items identify those processes which were able to maintain effluent concentrations below an upper limit value (total of 30).

Effluent Toxicity, Item 14

The effluent from the pilot plants (except ASP) were sampled on various occasions and tested for toxicity. Since the ASP did not produce sufficient volume, toxicity results from the Little River PCP will be used. This item will be ranked based on the percent of time the effluent was non-lethal to both rainbow trout and Daphnia magna. A weighing of 3 is proposed (total of 30).

A fifth category to access the pilot plants capability for producing an enhanced, potential tertiary quality effluent is being considered and has not been included yet.

HYDRAULIC PERFORMANCE (CATEGORY B)

A weighting of three is proposed for each of the items in this heading, since all are important to the design of a full-scale facility at the plant (for a total of 90).

Average Flow, Item 15

This item investigates which process was able to operate at the hydraulic loading which was initially determined for the process in Technical Memorandum No. 1. The average flow treated by the unit divided by the design flow rate times 10 is the ranking for each process.

Diurnal Flow, Item 16

The diurnal flow through each pilot plant was initiated to simulate the pattern of flow which the full-scale plant currently treats. The pilot plants were ranked based on the performance of each, either zero if the diurnal flow could not be handled, 5 if handling the diurnal flow pattern was a problem and 10 if the process could handle the diurnal flow pattern.

Peak Daily Flows, Item 17

Each pilot plant treated peak flows on a number of occasions, this item ranks whether the unit was able to treat a peak factor of 2.2 above the average flow. The full-scale plant is designed to handle a 2.2 peak factor. The ranking will be based on a ratio of the peak day flow to the average day flow, with 10 being the highest score.

SOLIDS HANDLING AND YIELD (CATEGORY C)

This heading summarizes the type and amount of solids generated by each process. These items will be ranked with the best process ranking 10, and each remaining process being ranked compared to the best one. The remaining processes score will be based on a ratio of its performance to the best process's performance times 10. Since sludge production and handling will be important to any full-scale plant, a weighting of three is proposed (for a total of 90).

Solids Yields, Item 18

This item ranks the solids yields of each process, based the amount of solids generated per organic load reduced. The unit producing the least solids per organic load reduced rank the highest.

Settleability of Solids, Item 19

Based on the results of the solids treatability task, the unit which had the best settling sludge based on the bench-top settling tests will rank a 10.

Dewaterability of Solids, Item 20

Based on the results of the solids treatability task, the unit which had the best dewatering sludge based on the bench-top dewatering tests will rank a 10.

OPERATIONAL CONCERNS (CATEGORY D)

Items 21 - 36

The fourth heading or category called operational concerns will include a number of items relating to the operation of each of the processes. These items will be ranked based on the pooled results of a questionnaire or ranking sheet completed by members of Steering Committee. Table 2 is a questionnaire for the items in this category, please complete the sheet and fax it to CH2M HILL by August 30, 1991 (fax no. (519)579-8986). To assist in ranking the proposed plant expansion layout, the preliminary layouts for each process are attached (Figures 1 through 4). The list of items proposed for this heading include:

- operational ease
- operational flexibility
- environmental concerns
- operational knowledge of system in Ontario or City of Windsor
- expected land requirements or "foot-print" of proposed plant
- effluent aesthetics
- ability of process to be constructed in stages
- startup requirements of the process
- ability to control the process
- automation potential of the process
- ability of process to handle seasonal loadings
- operational sampling requirements
- potential for operational problems
- potential for maintenance problems
- odour potential
- noise potential

The weighting factor on each of these items will vary depending on the items's expected importance; however, these items are expected to have a weighting factor of between one and two (based on a weighting of 1, total of 160).

ECONOMICS (CATEGORY E)

Items 37 - 39

The fifth heading entitled economics will rank the units based on the initial capital cost, yearly operational and maintenance costs, and the amortized value of the project over 20 years. This heading will not be completed until the costing has been completed by LaFontaine, Cowie, Buratto and Associates Ltd. The cost of the full-scale plant will be critical to the City; therefore a weighting factor of ten is proposed (for a total of 300).

The overall ranking will have a possible total of 1,000 points, with 360, 90, 90, 160, and 300 points proposed for each major category (see Table 1).

TABLE 1 — RANKING OF PILOT PLANT PERFORMANCE

PROCESS	CAT 'A' EFFLUENT PERFORMANCE														CAT 'B' HYD. PERF.			CAT 'C' SOLIDS PROD			CAT 'D' OPERATIONAL CONCERNS																CAT 'E' ECONOMICS			TOTALS					
	MONTHLY COMPLIANCE					PROB. LESS THAN OBJECTIVE					PROB LESS THAN UPPER																																		
	TSS	BOD5	TP	NH_3-N	UN-IONIZED NH_3-N	TSS	BOD5	TP	NH_3-N	UN-IONIZED NH_3-N	TSS	BOD5	TP	EFF. TOXICITY	AVG Q	DURNIAL Q	PEAK DAY Q	SOLIDS YIELD	SETTLE CHAR	DEWATER CHAR	OPER. EASE	OPER. FLEX	ENVIR. CONCERNS	SYSTEMS KNOW	PLANT SIZE	EFFUENT AESTHETIC	CONST IN STAGES	STARTUP EASE	CONTROL	AUTOMATION	SEASONAL LOADINGS	SAMPLING REQ'TS	OPER. PROBLEMS	MAINT. PROBLEMS	ODOURS	NOISE	CAPITAL COST	O&M COST	AMORTIZED COST	CATEGORY A	CATEGORY B	CATEGORY C	CATEGORY D	CATEGORY E	TOTAL
WEIGHTING 100 %	5 50	5 50	5 50	5 50	5 50	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	3 30	3 30	3 30	3 30	3 30	3 30	3 30	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	1 10	10 100	10 100	10 100	360	90	90	160	300	1000
TF/SC																																								0 0	0 0	0 0	0 0	0 0	0
RBC																																								0 0	0 0	0 0	0 0	0 0	0
BAF																																								0 0	0 0	0 0	0 0	0 0	0
ASP																																								0 0	0 0	0 0	0 0	0 0	0

Table 2
Questionnaire for Operational Concerns

Item	Processes *				Weighting
	TF/SC	RBC	BAF	ASP	
Operational ease					1
Operational flexibility					1
Environmental concerns					1
Operational Knowledge of System in Ontario or City of Windsor					1
Expected Land Requirements or "foot-print" of proposed plant					1
Effluent aesthetics					1
Ability of process to be constructed in stages					1
Startup requirements of the process					1
Ability to control the process					1
Automation potential of the process					1
Ability of process to handle seasonal loadings					1
Operational sampling requirements					1
Potential for operational problems					1
Potential for maintenance problems					1
Odour potential					1
Noise potential					1

Note: To be completed and faxed to CH2M HILL by August 30, 1991

* Each process to be ranked from 0 to 10, with 10 being the best.

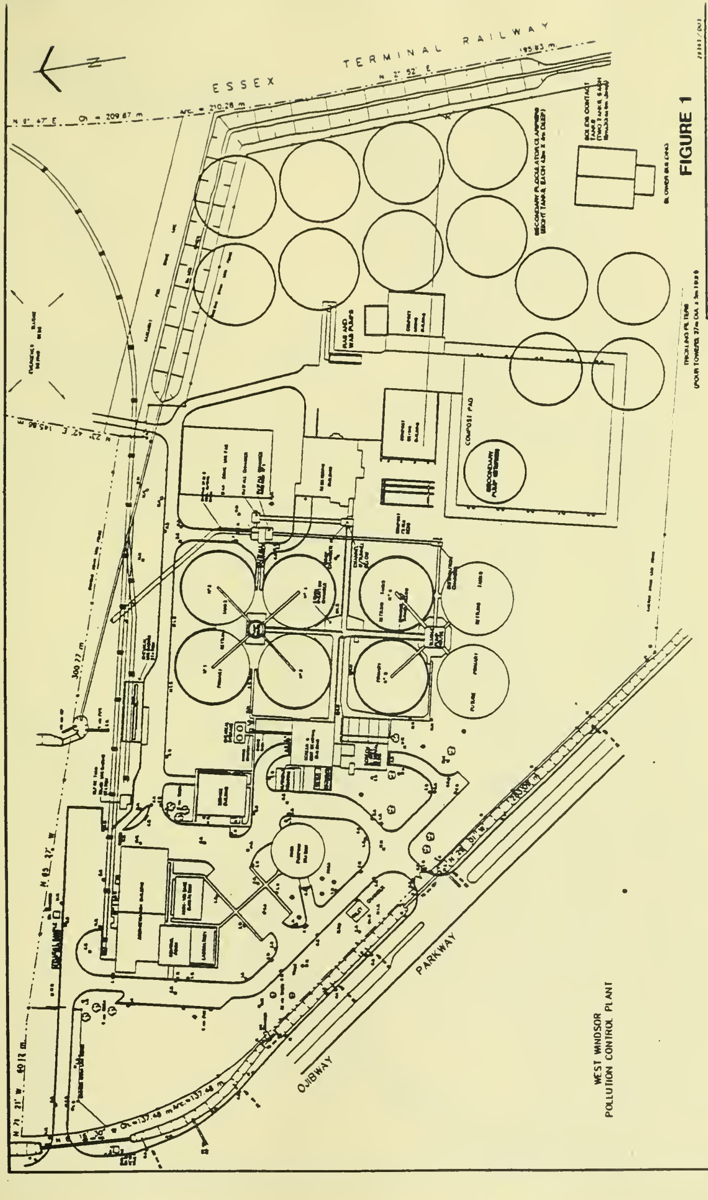

FIGURE 1

WEST WINDSOR
WATER POLLUTION CONTROL PLANT

POSSIBLE LAYOUT
FOR THE WEST WINDSOR PCP UPGRADE PROCESS
THICKENED SOLIDS CONTACT

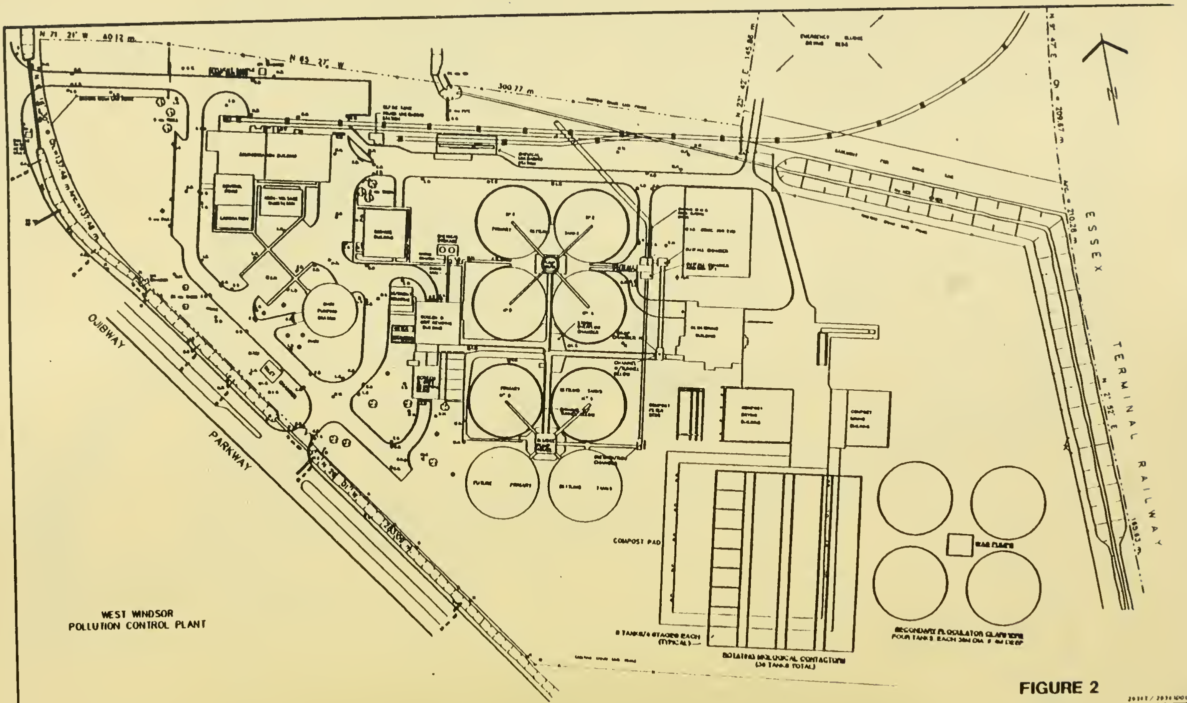

FIGURE 2

30101 / 3030000

WEST WINDSOR
WATER POLLUTION CONTROL PLANT

POSSIBLE LAYOUT
FOR THE WEST WINDSOR PCP UPGRADE PROCESS
ROTATING BIOLOGICAL CONTACTOR

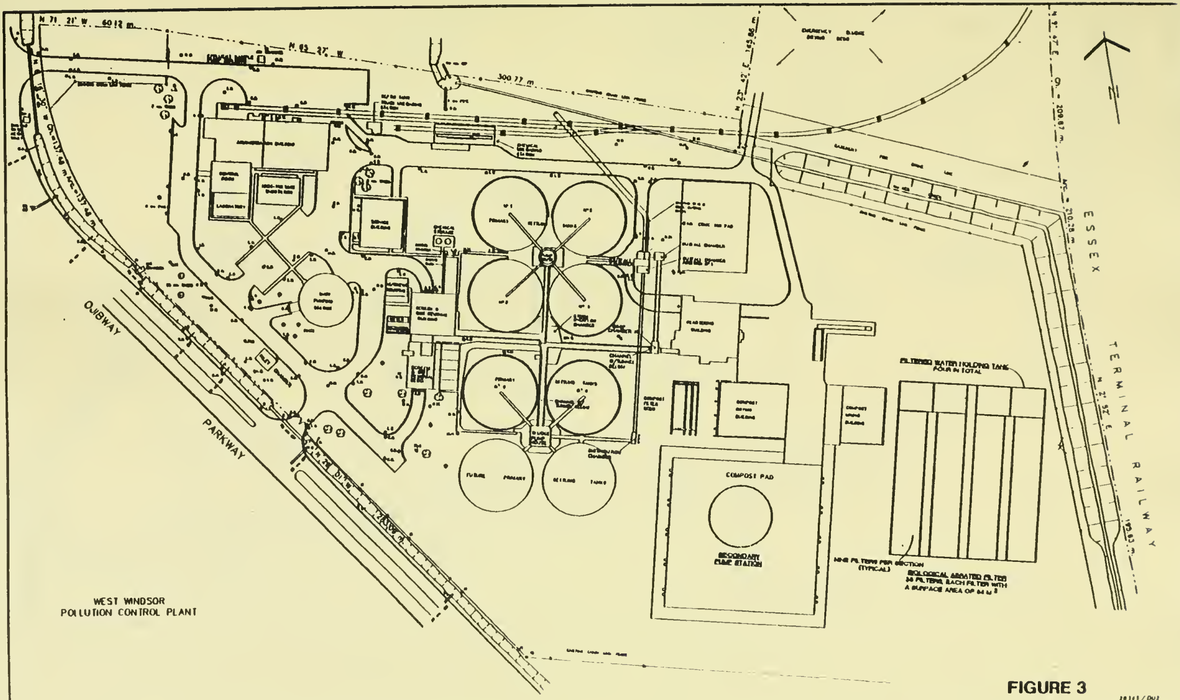

FIGURE 3

10161/001

WEST WINDSOR
WATER POLLUTION CONTROL PLANT

POSSIBLE LAYOUT
FOR THE WEST WINDSOR PCP UPGRADE PROCESS
BIOLOGICAL AERATED FILTER

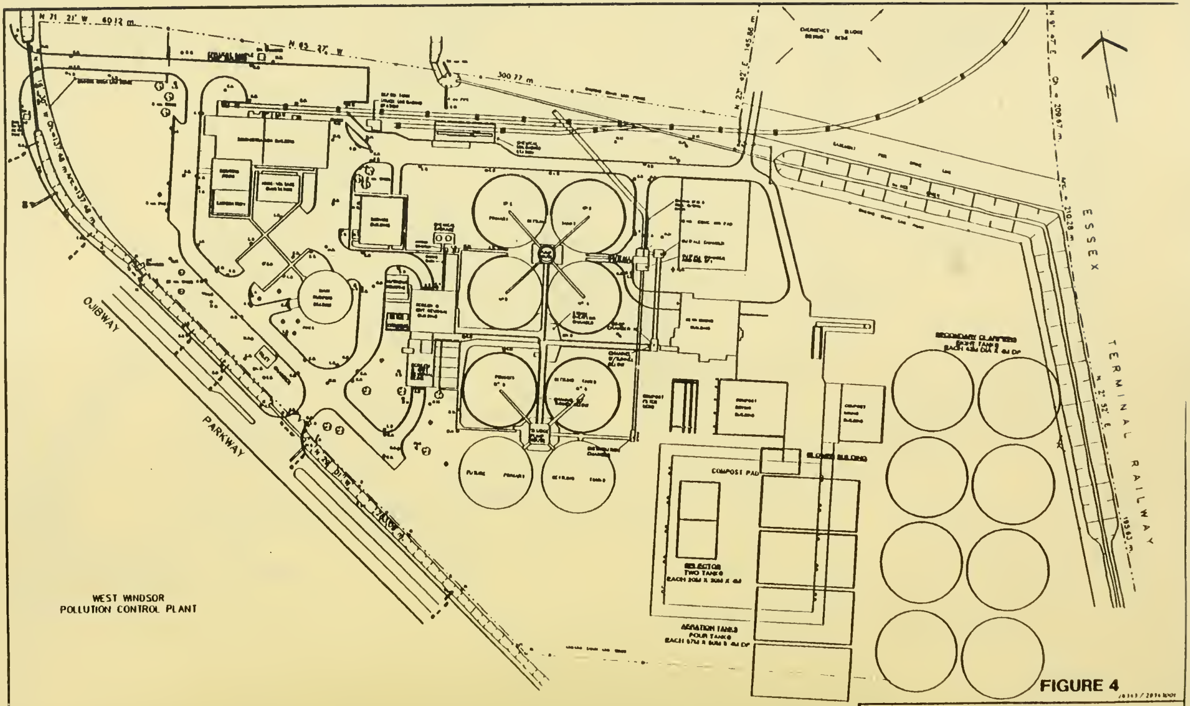

FIGURE 4

WEST WINDSOR
WATER POLLUTION CONTROL PLANT

POSSIBLE LAYOUT
WEST WINDSOR PCP UPGRADE PROCESS
[illegible] TRADITIONAL ACTIVATED SLUDGE

Appendix J

COMPARATIVE COST ESTIMATES

WEST WINDSOR POLLUTION CONTROL PLANT

INNOVATIVE TREATMENT ALTERNATIVES

FULL-SCALE PROCESS COST ESTIMATES

LA FONTAINE, COWIE, BURATTO & ASSOCIATES LIMITED

WEST WINDSOR POLLUTION CONTROL PLANT

FULL SCALE PROCESS COST ESTIMATES

BACKGROUND

This appendix to the report on "Evaluation of Innovative Alternatives for the West Windsor PCP Upgrade" contains only cost comparison information. Comments and recommendations on the comparitive merits of the alternative processes are contained in the Executive Summary Conclusions and Recommendations and in the body of the report.

The West Windsor Pollution Control Plant is located on a 14.6 hectare (36 acres) site at the corner of Ojibway Parkway and Sandwich Street. The original plant went into operation in 1970 as a primary treatment facility with the rated capacity of 109,000 m^3/d (24 MIGD). The level of treatment was upgraded to "physical - chemical" in 1973 to meet Provincial phosphorous removal requirements. The plant was expanded in 1980 to the present capacity of 159,000 m^3/d (35 MIGD).

The original plant layout was based on the ultimate provision of a 273,000 m^3/d (60 MIGD) conventional activated sludge treatment plant on the 14.6 hectare site. Existing raw sewage pumping, screening and grit removal facilities are sized for the ultimate plant capacity. Final settling tank sizing at the time of the original design was based on providing the same surface area as provided for the primary settling tanks. Using current design standards for activated sludge plants, the surface area of the final settling tanks will be approximately twice the surface area of the primary tanks. As a consequence, there is not sufficient space on the original site to accommodate the ultimate plant capacity and it will be necessary to locate some tankage on City owned property on the south side of Ojibway Parkway. For the purposes of this study, it is assumed final settling tanks will be located on the south side of the Parkway and biological treatment tankage on the original site.

FULL SCALE PLANT LAYOUTS

The following full scale treatment alternatives have been considered in this study.

- Conventional activated sludge
- Modified activated sludge

- Trickling filter/solids contact (TF/SC)
- Rotating biological contactor (RBC)
- Biological aerated filter (BAF)

Tankage sizing requirements for each of these alternatives are summarized in Tables 1 - 5 respectively.

Plant layouts for each of the alternatives are shown in Figures 1 - 5 respectively.

The following points should be noted with respect to tankage sizing information.

- All tankage sizes and configurations were selected for an ultimate plant capacity of 273,000 m^3/d. In some cases this results in tankage modules that are slightly oversized or undersized for the 159,000 m^3/d plant. The discrepancies in sizing are marginal and do not significantly effect cost estimates.
- The conventional activated sludge alternative is based on MOE Design Guidelines calling for 6 hours hydraulic retention time (HRT) in the aeration tanks at design dry weather flow (DWF).
- The modified activated sludge alternative has a 3 hour HRT in the aeration tanks at design DWF.
- Final settling tanks for both the conventional and modified activated sludge processes are in accordance with MOE Design Guidelines for plants with a nitrification requirement.
- Both activated sludge treatment alternatives include anoxic selector tanks. Approximately 3,000 m^3 of selector volume is included for the 159,000 m^3/d plant and 5,300 m^3 for the 273,000 m^3/d plant.
- The RBC treatment process description shows two alternatives for sizing of final settling tanks. One alternative is based on conventional MOE Design Guidelines and results in 13 - 45 m diameter tanks being required for a 273,000 m^3/d plant. By comparison, tankage sizing based on criteria from the innovative test program results in 6 - 45 m diameter tanks.

  The MOE has reviewed the test program findings and has indicated acceptance of a peak clarifier surface overflow rate of three times the MOE Design Guidelines would be contingent on a process audit to establish the validity of the criteria for a full scale plant. The design layout, therefore, shows both clarifier design alternative and provides sufficient space for tankage to MOE Guidelines.

- The TF/FC alternative includes solids contact tankage sized for a 48 minute HRT at design DWF. In fact, it is likely a good portion of the solids contact volume could be provided in aerated channels conveying mixed liquor to the final settling tanks.

- One additional primary clarifier is required for the BAF alternative in the ultimate plant layout to account for the 13% backwash recycle stream which will be returned to co-settle with the raw sewage in the primary settling tanks. For the initial capacity of 159,000 m^3/d, the 13% recycle stream will result in a peak surface overflow rate of 62.65 m^3/m^2 day on the existing 6 primary settling tanks.

- The original conventional activated sludge plant layout included smaller final settling tanks and incorporated all tankage on the original site. An hydraulic analysis has confirmed that, with efficient hydraulic design, gravity flow would have been possible through the original plant layout. However, using large final settling tanks as required by today's design standards makes it necessary to construct part of the tankage on City owned property south of Ojibway Parkway. This separation of tankage and less compact layout eliminates the possibility of gravity flow and results in the need for additional pumping facilities.

All treatment alternatives, with the exception of the BAF process, require pumping facilities. For the purposes of this study, it is assumed the pumping facilities will be positioned to pump primary effluent for subsequent gravity flow through the biological portion of the treatment process.

During final design, detailed hydraulic calculations and complete information on soil conditions should be evaluated to select the best position for the pumping facilities both in terms of capital and operating costs. Primary effluent pumping would appear to offer a capital cost advantage by elevating downstream tankage to take advantage of foundation conditions which are best near the ground surface and deteriorate with depth. Final effluent pumping may achieve some operating cost savings if hydraulic calculations show gravity flow is possible during low flow/low river level conditions.

Primary effluent pumping will definitely be required for the TF/SC alternative.

CAPITAL COST ESTIMATES

Capital cost estimates have been developed using information from a variety of sources including:

- Actual costs for construction of activated sludge treatment facilities at the Little River Pollution Control Plant in the City of Windsor during the period 1987 - 1991.
- Literature information on capital costs and cost comparisons for activated sludge, TF/SC, RBC and BAF treatment facility. Literature sources include:
    - A U.S. Army Corps of Engineers August 1988 publication titled "The Trickling Filter/Solids Contact Process: Application to Army Wastewater Plants".
    - A 1981 Report by Brown and Caldwell titled, "Engineering - Economic Comparison of the TF/SC Process to Conventional Technology".
    - A 1992 Associated Engineering/Brown and Caldwell/Reid Crowther Report for the Greater Vancouver Regional District - Annacis Island and Lulu Island Wastewater Treatment Plants.
    - A December 1988 paper by Dillon and Thomas titled, "A Pilot-Scale Evaluation of the 'Biocarbon Process' for the Treatment of Settled Sewage and for Tertiary Nitrification of Secondary Effluent".
    - An April 1990 Report by Montgomery Consulting Engineers - "Technology Assessment of the BAF".
- Cost information for construction of BAF facilities in Sherbrooke, Quebec and TF/SC facilities in Central Valley, Utah.
- Cost information from RBC and BAF equipment suppliers.

Estimated capital cost comparisons for the five (5) treatment alternatives are presented in Table 6 for the 159,000 m^3/d plant and in Table 7 for the 273,000 m^3/d plant. **THE FOLLOWING CONDITIONS AND QUALIFICATIONS MUST BE TAKEN INTO CONSIDERATION WHEN REVIEWING THESE COST COMPARISONS:**

- Facilities common to all alternatives are not included in the cost estimates. Examples of common elements include; chlorination - dechlorination facilities;

site work including access roads, drainage, and process sewers; upgrading and expansion of the existing plant electrical distribution and instrumentation and control systems; potable and process water facilities.

- No allowance is included in the costs for known poor soil conditions at the West Windsor site. Cost estimates for the activated sludge alternatives are based on Little River Pollution Control Plant construction costs where the soil bearing capacity is 3,000 lbs./ft.^{2} or greater. Soil bearing capacities at the West Windsor site are more typically 2,000 lbs./ft.^{2} or less and vary with depth and location. Special provisions are necessary at the West Windsor site to accommodate differential settlement.

- No costs are included for standby electric power generator facilities.

- No allowances is made for any changes in sludge handling, digestion or dewatering.

- The BAF sizing information and layout shows spare cells to allow for cell cleaning. The spare cells are not included in the cost estimate since other alternatives do not include costs for backup capacity.

- No allowance is made for the possible need for grease/oil removal prior to the BAF units. Indications are that yearly cleaning of the cells will suffice.

- Primary effluent pumping facilities are assumed with a 15 foot static lift for the activated sludge and RBC alternatives and a 20 foot static lift for the TF/SC alternative.

- No costs are included for administration building or laboratory facilities.

- In general, the cost estimates assume open tankage without covers. For the BAF alternative the filter surface area will be open with pipe galleries and service areas enclosed. For the RBC process uninsulated fiberglass covers are included for the RBC units.

- Land costs are not included in the estimates. Land area requirements are generally the same for all alternatives with the exception of the BAF process. The land area south of Ojibway Parkway will not be required for treatment facilities if the BAF process is selected.

- The cost estimates do not include any allowance for engineering, contingencies, taxes or interim financing.

- Capital cost estimates are in terms of 1992 dollars. The Engineering News Record Construction Cost Index has been used to relate estimates from other areas and previous years to a mid 1992 Toronto ENR-CCI value of 6730.

The cost estimates are useful for comparison purposes and generally agree with process comparisons from literature sources. Use of the estimates to predict actual construction costs for any of the processes is not recommended as a more detailed level of design would be required to develop costs taking into account site specific variables.

If all of the foregoing factors were taken into consideration and included in the cost estimate, the actual cost of construction of any of the process alternatives at West Windsor would be approximately 1.75 to 2 times the cost totals shown in Tables 6 and 7.

Accuracy of Cost

The level of confidence in capital cost estimates increases as a project progresses from a conceptual stage to a detailed design stage. Typically, the degree of accuracy at an early stage may be in the range of ± 30% or greater becoming more refined to a level of ± 10% at the end of the final design phase. Given the degree of detail possible in this study, it is estimated the level of accuracy of the capital cost figures presented in Tables 6 and 7 is in the range of ± 20%.

OPERATING AND MAINTENANCE COSTS

The main components that make up operating and maintenance costs for a pollution control plant include:

- Manpower costs
- Utility costs
    - Electrical power
    - Natural gas
    - Water
- Chemical costs
- Maintenance materials costs

Of these cost elements, only electrical energy costs have been used to develop a comparison of operating costs for the alternative treatment processes. The other cost elements have been excluded from the comparison for the following reasons:

- Operating and maintenance manpower costs are difficult to predict with any accuracy. On a qualitative basis, the evaluation of Operating Concerns presented in Chapter 5 indicates there is little difference in rating of the alternatives. The RBC process was classified as requiring the least operator attention. As a general observation, it would seem the more compact layout of the BAF process and the absence of a primary effluent pumping station and final settling tanks should result in less manpower requirements.
- Natural gas and municipal water usage are expected to be generally the same for all alternatives.
- The cost of polymer and iron or aluminum salts for phosphorous removal in the primary clarifiers will be the same for all alternatives as will the cost of chemicals for chlorination - dechlorination. Some additional chemical costs will be required for cleaning of the BAF cells but this cost will be partially offset by chemicals used for cleaning of final settling tank weirs and launders. It is expected the overall difference in chemical costs for the alternative processes will be minimal.
- Maintenance material costs are expected to be similar for all processes.

Horsepower requirements for each of the processes have been calculated and are presented in Tables 9 and 10. The following notes will assist in reviewing these tables.

- Installed horsepower figures are based on satisfying peak flow pumping and air delivery requirements and include a standby unit or units so that peak requirements can be met with the largest pump/blower out of operation.
- Running horsepower figures represent requirements under design DWF flow and loading conditions.
- Running horsepower requirements for primary effluent pumping in the TF/SC process assume an additional 5 feet of static lift as compared to the activated sludge and RBC alternatives. Also, an allowance is included in the TF/SC process for recycling effluent from the trickling filter to maintain a desirable wetting rate during low flow periods.

- Process air requirements and blower sizing will be similar for the activated sludge processes and the BAF process since BOD_5 and ammonia loadings to these processes will be identical.

- Running horsepower for the return activated sludge pumps used with the conventional and modified activated sludge processes has been calculated based on a typical return rate of 75% to 80%. A return rate of 40% to 50% has been assumed for the TF/SC process.

- Backwash water pumps for the BAF process are sized to provide a backwash flow rate of 40 $m^3/m^2/h$ and the total daily volume of backwash water is assumed to be 13% of the average daily flow.

- Backwash air blowers for the BAF process are sized to provide a 70 $m^3/m^2/h$ air flow rate for 40 minutes per day for each filter cell.

- Each RBC unit is equipped with a 5 horsepower motor drawing approximately 4.5 horsepower.

- Each clarifier is equipped with a 5 horsepower motor drawing approximately 3.75 horsepower.

Table 10 provides a comparison of annual electrical energy costs assuming a 1992 cost of 5¢ per kilowatt hour. This table shows the activated sludge processes have the highest electrical operating costs with the TF/SC process being the least expensive.

PRESENT VALUE ANALYSIS[1] [2]

In order to compare the overall economics of the different treatment processes, cash flow for both capital and operating costs must be converted to one point in time. Conversion of all cash flow to the present time is termed the present value or present worth of the process. The present value of any of the processes can be defined as the amount of money that would have to be invested at the outset (initial year) to repay the cost of the facility and pay for all operating, maintenance and miscellaneous costs over the design life of the facility. The present value investment would accrue interest while being used to pay for costs in inflated dollars.

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[1] Economic Analysis for Engineers and Managers
J. C. Sprague/J. D. Whittaker

[2] ABR Consultants Report for Annacis Island and Lulu Island Wastewater Treatment Plants

Determination of present value involves calculating all future costs over a selected design period in terms of current dollars and a discount rate is applied to reflect the proportional difference between the interest rate and inflation rate.

The Discount Rate (DR) is defined as:

$$DR = \frac{(1 + \text{Rate of Interest})}{(1 + \text{Rate of Inflation})} - 1$$

Tables 11 and 12 provide a Present Value Cost Comparison for each of the five processes over a 20 year period. An interest rate of 10% and an inflation rate of 4.5% have been used. An inflation rate of 6% has been selected for electrical power supply based on the reasoning that annual increases in the cost of electrical energy will be somewhat higher than the general rate of inflation.

The present value figures for both the 159,000 m^3/d and 273,000 m^3/d plants indicate the conventional activated sludge is the most expensive process. The modified activated sludge process appears to be marginally less expensive than the other innovative processes.

OBSERVATIONS

Given the ±20% level of accuracy in the capital cost estimates, the following general observations can be made.

- The capital cost and present value comparisons indicate there is a significant savings in using any one of the innovative treatment processes as compared to the conventional activated sludge process.
- Capital cost estimates for the 159,000 m^3/day plant (Table 6) show cost savings ranging from 13% to 24% using the innovative treatment alternatives with the modified activated sludge process being the least expensive.
- Present value comparisons for the 159,000 m^3/day plant (Table 11) indicate the costs of the four innovative processes are 12 to 20% less than the conventional process.
- Capital cost estimates for the 273,000 m^3/day plant (Table 7) show costs for the innovative processes are 11% to 24% less than for the conventional process with modified activated sludge being the least expensive.
- Present value comparisons for the 273,000 m^3/day plant (Table 12) show cost savings ranging from 11% to 21% using the innovative treatment processes.

The capital cost estimate tables show possible cost reductions for the RBC alternative if reduced final clarifier sizing is permitted. For the 159,000 m^3/day plant, the capital cost of the RBC alternative with reduced final clarifier sizing is estimated to be approximately 28% less than a conventional plant. For the 273,000 m^3/day plant there is an estimated 23% capital cost reduction. The present value analysis does not take the possible savings for smaller clarifiers into account. It is unclear how much emphasis should be placed on this possible savings in selecting a treatment process when, in fact, there is no assurance that larger clarifiers will not be required.

A 19 acre parcel of City owned land on the south side of Ojibway Parkway is required to accommodate final settling tankage and associated facilities for all alternatives except the BAF process. If this land were valued at $50,000.00 per acre which is a very reasonable cost for serviced industrial property, there would be a cost advantage of approximately $1,000,000.00 to the BAF alternative.

TABLE 1

WEST WINDSOR POLLUTION CONTROL PLANT

TREATMENT PROCESS: CONVENTIONAL ACTIVATED SLUDGE[4]

UNIT PROCESS	INITIAL CAPACITY 159,000 m^3/d DWF[1] 350,000 m^3/d WWF[2]	ULTIMATE CAPACITY 273,000 m^3/d DWF 600,000 m^3/d WWF
Preliminary Treatment (Screening & Grit Removal)	Existing	Existing
Primary Settling Tanks[3]	- 6 - 36.6 m dia. Existing Clarithickner Tanks	- 10 - 36.6 m dia. Clarithickener Tanks
Aeration Tanks[4]	- 39,750 m^3 Volume - 7,360 m^2 Surface Area at 5.4 m depth - 12 Tanks each 10 m wide and 64 m long	- 68,250 m^3 Volume - 12,640 m^2 Surface Area at 5.4 m depth - 20 Tanks each 10 m wide and 64 m long
Final Settling Tanks[5]	- 12,720 m^2 Area Based on Solids Loading Rate - 8 - 45 m dia. Tanks with 5 m SWD (12,723 m^2 Area)	- 21,800 m^2 Based on Solids Loading Rate - 14 - 45 m dia. Tanks with 5 m SWD (22,266 m^2 Area)
Disinfection[6]	- 3,300 m^3 Volume - 1 Tank 5 m deep, 10 m wide & 57 m long (Enlarge when DWF > 136,500 m^3/D)	- 5,700 m^3 Volume - 2 Tanks 5 m deep, 10 m wide & 57 m long

1) DWF = Dry Weather Flow
2) WWF = Wet Weather Flow = 2.2 x DWF
3) Primary Tanks SOR = 25.95 m^3/m^2/Day at DWF; 57.09 m^3/m^2 Day at WWF
4) Aeration Tank HRT = 6 hrs. at DWF per MOE Design Guidelines
5) Peak SOR of 29.4 m^3/m^2 Day for Nitrification.
Solids Loading Rate (SLR) no greater than 120 $kg/m^2/d$ at peak flow including 100% DWF return sludge
6) HRT of 30 min at DWF. Dechlorination facilities may also be required.

TABLE 2

WEST WINDSOR POLLUTION CONTROL PLANT
TREATMENT PROCESS: MODIFIED ACTIVATED SLUDGE[4]

UNIT PROCESS	INITIAL CAPACITY 159,000 m^3/d DWF[1] 350,000 m^3/d WWF[2]	ULTIMATE CAPACITY 273,000 m^3/d DWF 600,000 m^3/d WWF
Preliminary Treatment (Screening & Grit Removal)	Existing	Existing
Primary Settling Tanks[3]	- 6 - 36.6 m dia. Existing Clarithickner Tanks	- 10 - 36.6 m dia. Clarithickener Tanks
Aeration Tanks[4]	- 19,875 m^3 Volume - 3,680 m^2 Surface Area t 5.4 m depth - 6 Tanks each 10 m wide and 64 m long	- 34,125 m^3 Volume - 6,320 m^2 Surface Area at 5.4 m depth - 10 Tanks each 10 m wide and 64 m long
Final Settling Tanks[5]	- 12,720 m^2 Area Based on Solids Loading Rate - 8 - 45 m dia. Tanks with 5 m SWD (12,723 m^2 Area)	- 21,800 m^2 Based on Solids Loading Rate - 14 - 45 m dia. Tanks with 5 m SWD (22,266 m^2 Area)
Disinfection[6]	- 3,300 m^3 Volume - 1 Tank 5 m deep, 10 m wide & 57 m long (Enlarge when DWF > 136,500 m^3/D)	- 5,700 m^3 Volume - 2 Tanks 5 m deep, 10 m wide & 57 m long

1) DWF = Dry Weather Flow
2) WWF = Wet Weather Flow = 2.2 x DWF
3) Primary Tanks SOR = 25.95 m^3/m^2/Day at DWF
= 57.09 m^3/m^2 Day at WWF
4) Aeration Tank HRT = 3 hrs. at DWF
5) Peak SOR of 29.4 m^3/m^2 Day for Nitrification Application.
Solids Loading Rate (SLR) no greater than 120 $kg/m^2/d$ peak flow including 100% DWF return sludge.
6) HRT of 30 min. at DWF. Facilities for dechlorination may also be required.

TABLE 3

WEST WINDSOR POLLUTION CONTROL PLANT

TREATMENT PROCESS: ROTATING BIOLOGICAL CONTACTOR (RBC)

UNIT PROCESS	INITIAL CAPACITY 159,000 m^3/d DWF[1] 350,000 m^3/d WWF[2]	ULTIMATE CAPACITY 273,000 m^3/d DWF 600,000 m^3/d WWF
Preliminary Treatment (Screening & Grit Removal)	Existing	Existing
Primary Settling Tanks[3]	- 6 - 36.6 m dia. Existing Clarithickner Tanks	- 10 - 36.6 m dia. Clarithickener Tanks
Rotating Biological Contactors	- 2.16×10^6 m^2 Media Surface Area Required - 16 Trains of 10 shafts per train and 13,975 m^2 per shaft (High Density) (Surface Area $= 2.24 \times 10^6 m^2$) - Tankage footprint approx. 82 m x 100 m	- 3.7 $10^6 m^2$ Media Surface Area Required - 27 Trains of 10 shafts/ train and 13,975 m^2/ shaft (Surface Area $= 3.77 \times 10^6 m^2$) - Tankage footprint approx. 146 m x 100 m
Final Settling Tanks		
A) Conventional MOE Criteria SOR = 29.4 m^3/m^2 Day at Peak Flow	- 11,900 m^2 Area - 8 - 45 m dia. Tanks	- 20,410 m^2 Area - 13 - 45 m dia. Tanks
B) Criteria From Innovative Test Programme SOR = 31 m^3/m^2 Day at DWF and 75 m^3/m^2 Day at WWF	- 5,129 m^2 area - 4 - 45 m dia. Tanks (6,362 m^2 Area)	- 8,806 m^2 Area - 6 - 45 m dia. Tanks (9,543 m^2 Area)
Disinfection[4]	- 3,300 m^3 Volume - 1 Tank 5 m deep, 10 m wide & 57 m long (Enlarge when DWF > 136,500 m^3/D)	- 5,700 m^3 Volume - 2 Tanks 5 m deep, 10 m wide & 57 m long

1) DWF = Dry Weather Flow
2) WWF = Wet Weather Flow = 2.2 x DWF
3) Primary Tanks SOR = 25.95 m^3/m^2/Day at DWF
= 57.09 m^3/m^2 Day at WWF
4) HRT of 30 min. at DWF. Facilities for dechlorination may also be required.

TABLE 4

WEST WINDSOR POLLUTION CONTROL PLANT
TREATMENT PROCESS: TRICKLING FILTER/SOLIDS CONTACT (TF/SC)

UNIT PROCESS	INITIAL CAPACITY 159,000 m^3/d DWF[1] 350,000 m^3/d WWF[2]	ULTIMATE CAPACITY 273,000 m^3/d DWF 600,000 m^3/d WWF
Preliminary Treatment (Screening & Grit Removal)	Existing	Existing
Primary Settling Tanks[3]	- 6 - 36.6 m dia. Existing Clarithickner Tanks	- 10 - 36.6 m dia. Clarithickener Tanks
Trickling Filter Tanks	- 4,540 m^2 Surface Area Required (Wetting Rate of 35 $m^3/m^2/d$) - Five Metre Tower Depth - 3 - 45 m dia. Tanks with 5 m media depth	- Five Metre Tower Depth - 5 - 45 m dia. Tanks with 5 m media depth
Solids Contact Tanks[4]	- 3 Tanks each 15 m x 24 m x 5 m deep	-6 Tanks each 15 m x 24 m x 5 m deep
Final Settling Tanks Conventional MOE Criteria SOR = 29.4 m^3/m^2 Day at Peak Flow	- 11,900 m^2 Area - 8 - 45 m dia. Tanks	- 20,410 m^2 Area - 13 - 45 m dia. Tanks
Disinfection[5]	- 3,300 m^3 Volume - 1 Tank 5 m deep, 10 m wide & 57 m long (Enlarge when DWF > 136,500 m^3/D)	- 5,700 m^3 Volume - 2 Tanks 5 m deep, 10 m wide & 57 m long

1) DWF = Dry Weather Flow
2) WWF = Wet Weather Flow = 2.2 x DWF
3) Primary Tanks SOR = 25.95 m^3/m^2/Day at DWF
= 57.09 m^3/m^2 Day at WWF
4) Scale up from pilot plant operations. Provides 48 min HRT at Design Flow and 21.8 min. at peak flow without recycle.
5) HRT of 30 min. at DWF - Facilities for dechlorination may also be required.
6) Information from the Central Valley TF/SC treatment facility in Salt Lake City, Utah, indicates the use of an improved design flocculant clarifier may achieve a reduction in final clarifier sizing similar to that shown for the RBC process.

TABLE 5

WEST WINDSOR POLLUTION CONTROL PLANT

TREATMENT PROCESS: BIOLOGICAL AERATED FILTER (BAF)

UNIT PROCESS	INITIAL CAPACITY 159,000 m^3/d DWF[1] 350,000 m^3/d WWF[2] 395,500 m^3/d Peak Flow[3]	ULTIMATE CAPACITY 273,000 m^3/d DWF 600,000 m^3/d WWF 678,000 m^3/d Peak Flow
Preliminary Treatment (Screening & Grit Removal)	Existing	Existing
Primary Settling Tanks[4]	- 6 - 36.6 m dia. Existing Clarithickner Tanks	- 11 - 36.6 m dia. Clarithickener Tanks
BAF Units[5]	- 2,996 m^2 Surface Area Required - 36 Cells required each 84 m^2 area & 2.5 m media depth - Use 40 Cells to account for cells being out of service for cleaning	- 5,136 m^2 Surface Area Required - 62 Cells required each 84 m^2 area & 2.5 m media depth - Use 68 Cells to account for cells being out of service for cleaning
Final Settling Tanks	None Required	None Required
Disinfection[6]	- 3,300 m^3 Volume - 1 Tank 5 m deep, 10 m wide & 57 m long	- 5,700 m^3 Volume - 2 Tanks 5 m deep, 10 m wide & 57 m long

1) DWF = Dry Weather Flow
2) WWF = Wet Weather Flow = 2.2 x DWF
3) Peak Flow = 1.13 x WWF to account for 13% backwash recycle stream
4) Primary settling tank design based on peak flow figure gives a SOR of 62.65 m^3/m^2 day for the initial capacity and 58.58 m^3/m^2 day for ultimate capacity.
5) Filter cell sizing based on peak flow capability of 5.5 m^3/m^2 hour.
6) HRT of 30 min. at DWF; dechlorination facilities may also be required.

TABLE 6
WEST WINDSOR POLLUTION CONTROL PLANT
CAPITAL COST COMPARISON: (1),(2) 159,000 M^3/DAY PLANT CAPACITY

Component/Unit Process	Conventional Activated Sludge	Modified Activated Sludge	Trickling Filter/Solids Contact	Rotating Biological Contactor	Biological Aerated Filter
Raw Sewage Pumping, Screening, Grit Removal	Existing	Existing	Existing	Existing	Existing
Primary Settling Tanks	Existing	Existing	Existing	Existing	Existing(3)
Primary Effluent Pumping Station	$ 1,800,000.	$ 1,800,000.	$ 1,800,000.	$ 1,800,000.	N/A
Biological Process Tankage and Equipment	$19,160,000.	$ 9,590,000.	$16,250,000.	$18,590,000.	$32,960,000.
Blower Building & Equipment	$ 2,280,000.	$ 2,280,000.	$ 900,000.	N/A	Included Above
Return/Waste Activated Sludge Pump Building & Equipment	$ 1,980,000.	$ 1,980,000.	$ 990,000.	N/A	N/A
Final Settling Tanks	$11,260,000.	$11,260,000.	$11,260,000.(5)	$11,260,000. (4) ($5,630,000.)	N/A
Chlorination/Dechlorination	Common to All Process Alternatives				
New Outfall	$ 1,450,000.	$ 1,450,000.	$ 1,450,000.	$ 1,450,000.	N/A
Pipe and Channel Connections	$ 1,230,000.	$ 1,230,000.	$ 1,230,000.	$ 1,230,000.	N/A
RAS/WAS Piping/Channels	$ 500,000.	$ 500,000.	$ 500,000.	N/A	N/A
TOTAL (1)	$39,860,000.	$30,290,000.	$34,380,000.	$34,330,000. ($28,700,000.)(4)	$32,960,000.

1) **Costs are for comparison purposes and do not represent total construction costs. Refer to text of report for details.**
2) Cost estimates are in terms of 1992 dollars.
3) 13% backwash water recycle will theoretically hydraulically overload primary settling tanks during peak flow period. The tanks can tolerate this loading without significant impairment of effluent quality.
4) Eight (8) clarifiers are required to maintain a conventional surface overflow rate (SOR) of 29.4 m^3/m^2 day at peak flow. Only four (4) clarifiers are required if a peak SOR of 75 m^3/m^2 day is permitted in accordance with innovative testing results
5) Information from the Central Valley TF/SC treatment facility in Salt Lake City, Utah indicates the use of an improved design flocculant clarifier may achieve a reduction in final clarifier sizing and cost savings similar to that shown for the RBC process.

TABLE 7
WEST WINDSOR POLLUTION CONTROL PLANT
CAPITAL COST COMPARISON: [1],[2] 273,000 M³/DAY PLANT CAPACITY

Component/Unit Process	Conventional Activated Sludge	Modified Activated Sludge	Trickling Filter/Solids Contact	Rotating Biological Contactor	Biological Aerated Filter
Raw Sewage Pumping, Screening, Grit Removal	Existing	Existing	Existing	Existing	Existing
Primary Settling Tanks	$ 6,200,000.	$ 6,200,000.	$ 6,200,000.	$ 6,200,000.	$ 7,750,000.
Primary Effluent Pumping Station	$ 2,100,000.	$ 2,100,000.	$ 2,100,000.	$ 2,100,000.	N/A
Biological Process Tankage and Equipment	$32,840,000.	$16,420,000.	$27,860,000.	$31,910,000.	$51,360,000.
Blower Building & Equipment	$ 2,600,000.	$ 2,600,000.	$ 1,300,000.	N/A	Included Above
Return/Waste Activated Sludge Pump Building & Equipment	$ 2,180,000.	$ 2,180,000.	$ 1,090,000.	N/A	N/A
Final Settling Tanks	$18,340,000.	$18,340,000.	$17,030,000.[4]	$17,030,000. [3] ($7,860,000.)	N/A
Chlorination/Dechlorination	Common to All Process Alternatives				
New Outfall	$ 1,450,000.	$ 1,450,000.	$ 1,450,000.	$ 1,450,000.	N/A
Pipe and Channel Connections	$ 1,230,000.	$ 1,230,000.	$ 1,230,000.	$ 1,230,000.	N/A
RAS/WAS Piping/Channels	$ 500,000.	$ 500,000.	$ 500,000.	N/A	N/A
TOTAL [1]	$67,440,000.	$51,020,000.	$58,760,000.	$59,920,000. ($50,750,000.)[3]	$59,920,000.

1) Costs are for comparison purposes and do not represent total construction costs. Refer to text of report for details.
2) Cost estimates are in terms of 1992 dollars.
3) Thirteen (13) clarifiers are required for conventional surface overflow rate (SOR) of 29.4 m^3/m^2 day at peak flow. Only six (6) clarifiers are required if a peak SOR of 75 m^3/m^2 day is used in accordance with innovative testing results.
4) Information from the Central Valley TF/SC treatment facility in Salt Lake City, Utah indicates the use of an improved design flocculant clarifier may achieve a reduction in final clarifier sizing and cost savings similar to that shown for the RBC process.

TABLE 8

WEST WINDSOR POLLUTION CONTROL PLANT - SUMMARY OF HORSEPOWER REQUIREMENTS

273,000 M^3/DAY PLANT

Process	Activated(1) Sludge	Trickling Filter/Solids Contact	Rotating Biological Contactor	Biological Aerated Filter
Primary Settling Tanks	N/A	N/A	N/A	5 HP installed 4 HP running
Primary Effluent Pumping	1,000 HP installed(2) 275 HP running(3)	1,000 HP installed 485 HP running(4)	1,000 HP installed 275 HP running	N/A
Process Air Blowers	3,000 HP installed 1065 HP running	750 HP installed 300 HP running	N/A	3,000 HP installed 1,065 HP running
Return Activated Sludge Pumps	600 HP installed 350 HP running	400 HP installed 150 HP running	N/A	N/A
Backwash Water Pumps	N/A	N/A	N/A	600 HP installed 150 HP - 12 hrs/day (Average 75 HP running)
Backwash Air Blower	N/A	N/A	N/A	600 HP installed 300 HP - 20 hrs/day (Average 250 HP running)
RBC Drives	N/A	N/A	1,350 HP installed 1,215 HP running	N/A
Final Settling Tanks	70 HP installed 50 HP running	65 HP installed 50 HP running	65 HP installed 50 HP running	N/A
SUMMARY	**4,670 HP installed 1,740 HP running**	**2,215 HP installed 985 HP running**	**2,415 HP installed 1,540 HP running**	**4,205 HP installed 1,394 HP running**

1) Conventional and modified activated sludge are assumed to have identical energy requirements.
2) Installed HP includes standby units
3) Running HP under average design conditions
4) Includes an allowance for effluent recycle to maintain the trickling filter wetting rate during low diurnal flow periods.

TABLE 9

WEST WINDSOR POLLUTION CONTROL PLANT - SUMMARY OF HORSEPOWER REQUIREMENTS

159,000 M^3/DAY PLANT

Process	Activated(1) Sludge	Trickling Filter/Solids Contact	Rotating Biological Contactor	Biological Aerated Filter
Primary Settling Tanks	Commom to All Alternatives			
Primary Effluent Pumping	600 HP installed(2) 160 HP running(3)	600 HP installed 285 HP running(4)	600 HP installed 160 HP running	N/A
Process Air Blowers	1,800 HP installed 550 HP running	450 HP installed 150 HP running	N/A	1,800 HP installed 550 HP running
Return Activated Sludge Pumps	450 HP installed 200 HP running	300 HP installed 100 HP running	N/A	N/A
Backwash Water Pumps	N/A	N/A	N/A	450 HP installed 150 HP - 7 hrs/day (Average 45 HP running)
Backwash Air Blower	N/A	N/A	N/A	450 HP installed 150 HP - 24 hrs/day
RBC Drives	N/A	N/A	800 HP installed 720 HP running	N/A
Final Settling Tanks	40 HP installed 30 HP running	40 HP installed 30 HP running	40 HP installed 30 HP running	N/A
SUMMARY	**2,890 HP installed 940 HP running**	**1,390 HP installed 565 HP running**	**1,440 HP installed 910 HP running**	**2,700 HP installed 745 HP running**

1) Conventional and modified activated sludge are assumed to have identical energy requirements.
2) Installed HP includes standby units
3) Running HP under average design conditions
4) Includes an allowance for effluent recycle to maintain the trickling filter wetting rate during low diurnal flow periods.

TABLE 10

WEST WINDSOR POLLUTION CONTROL PLANT

ESTIMATED ANNUAL ELECTRICAL ENERGY COSTS(1)

Design Capacity	Activated Sludge	Trickling Filter/Solids Contact	Rotating Biological Contactor	Biological Aerated Filter
159,000 m^3/day	$307,000.	$185,000.	$297,000.	$243,000.
273,000 m^3/day	$568,000.	$322,000.	$503,000.	$455,000.

1) Electrical energy cost assumed to be 5¢/KWH in 1992

TABLE 11
West Windsor Pollution Control Plant
Capital Cost Comparison - Present Value Analysis
159 000 m3/Day Plant Capacity

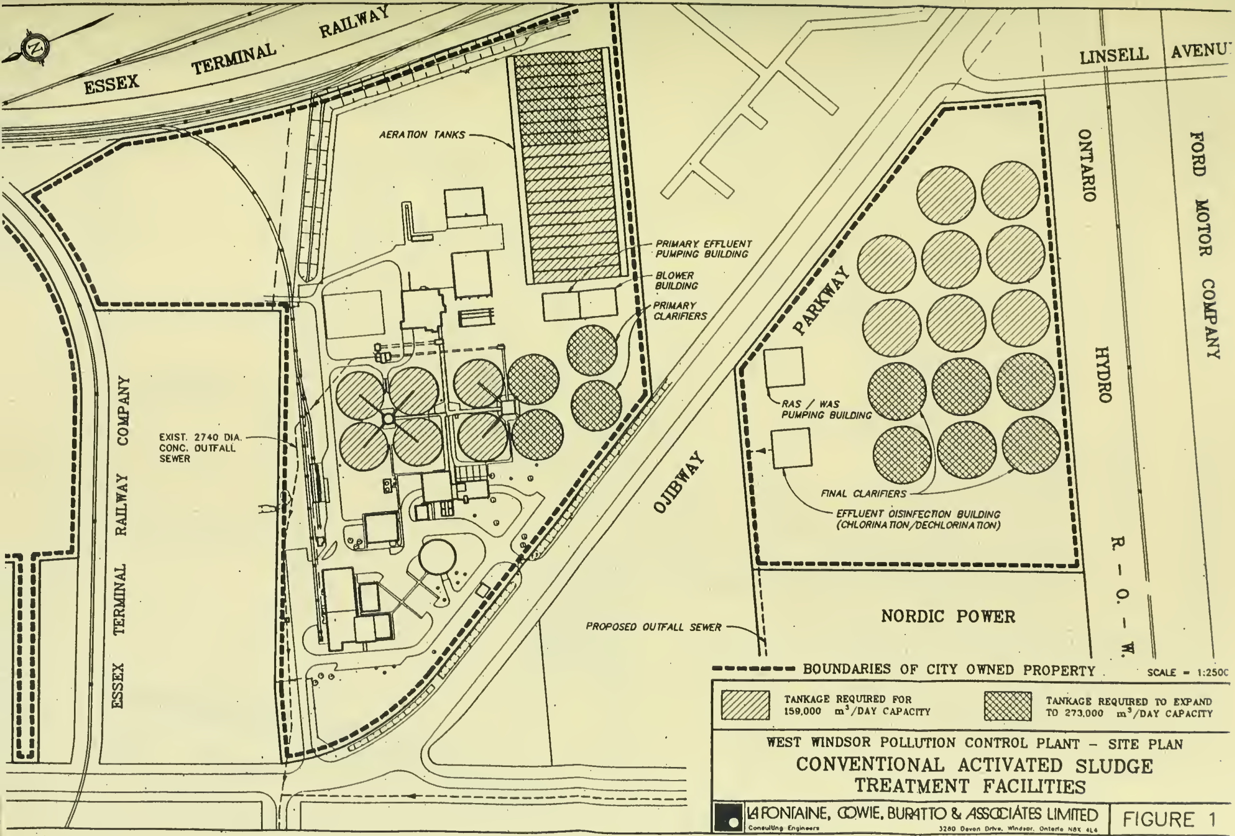
		Conventional Activated Sludge		Modified Activated Sludge		Trickling Filter Solids Contact		Rotating Biological Contactor		Biological Aerated Filter	
		Capital	Operating Cost	Capital	Operating Cost	Capital	Operating Cost	Capital	Operating Cost	Capital	Operating Cost
Capital Cost		$39,660,000		$30,290,000		$34,380,000		$34,330,000		$32,960,000	
Annual Payment	Year	$4,658,449	$307,000	$3,557,852	$307,000	$4,038,262	$185,000	$4,032,389	$297,000	$3,871,469	$243,000
Present Value	1	$4,425,526	$313,587	$3,379,959	$313,587	$3,836,349	$188,969	$3,830,769	$303,372	$3,677,896	$248,213
	2	$4,204,250	$320,314	$3,210,961	$320,314	$3,644,531	$193,023	$3,639,231	$309,881	$3,494,001	$253,539
	3	$3,994,037	$327,187	$3,050,413	$327,187	$3,462,305	$197,165	$3,457,269	$316,529	$3,319,301	$258,978
	4	$3,794,336	$334,206	$2,897,893	$334,206	$3,289,190	$201,395	$3,284,406	$323,320	$3,153,336	$264,535
	5	$3,604,619	$341,376	$2,752,998	$341,376	$3,124,730	$205,715	$3,120,186	$330,257	$2,995,669	$270,210
	6	$3,424,388	$348,701	$2,615,348	$348,701	$2,968,494	$210,129	$2,964,176	$337,342	$2,845,886	$276,007
	7	$3,253,168	$356,182	$2,484,581	$356,182	$2,820,069	$214,637	$2,815,968	$344,580	$2,703,591	$281,929
	8	$3,090,510	$363,823	$2,360,352	$363,823	$2,679,065	$219,242	$2,675,169	$351,973	$2,568,412	$287,978
	9	$2,935,985	$371,629	$2,242,334	$371,629	$2,545,112	$223,946	$2,541,411	$359,524	$2,439,991	$294,156
	10	$2,789,185	$379,602	$2,130,217	$379,602	$2,417,857	$228,751	$2,414,340	$367,237	$2,317,992	$300,467
	11	$2,649,726	$387,746	$2,023,707	$387,746	$2,296,964	$233,658	$2,293,623	$375,116	$2,202,092	$306,913
	12	$2,517,240	$396,065	$1,922,521	$396,065	$2,182,116	$238,671	$2,178,942	$383,164	$2,091,987	$313,498
	13	$2,391,378	$404,563	$1,826,395	$404,563	$2,073,010	$243,792	$2,069,995	$391,385	$1,987,388	$320,224
	14	$2,271,809	$413,243	$1,735,075	$413,243	$1,969,359	$249,022	$1,966,495	$399,782	$1,888,019	$327,094
	15	$2,158,218	$422,108	$1,648,322	$422,108	$1,870,891	$254,365	$1,868,170	$408,359	$1,793,618	$334,112
	16	$2,050,308	$431,165	$1,565,906	$431,165	$1,777,347	$259,822	$1,774,762	$417,120	$1,703,937	$341,280
	17	$1,947,792	$440,415	$1,487,610	$440,415	$1,688,479	$265,397	$1,686,024	$426,069	$1,618,740	$348,602
	18	$1,850,403	$449,864	$1,413,230	$449,864	$1,604,055	$271,091	$1,601,723	$435,210	$1,537,803	$356,081
	19	$1,757,882	$459,516	$1,342,568	$459,516	$1,523,853	$276,907	$1,521,636	$444,548	$1,460,913	$363,721
	20	$1,669,988	$469,374	$1,275,440	$469,374	$1,447,660	$282,848	$1,445,555	$454,085	$1,387,867	$371,524
Net Present Value		$56,780,748	$7,730,667	$43,365,831	$7,730,667	$49,221,435	$4,658,545	$49,149,851	$7,478,853	$47,188,438	$6,119,062
			$64,511,415		$51,096,498		$53,879,980		$56,628,704		$53,307,500

Interest Rate	10.00%	Inflation Rate	4.50%	Discount Rate (capital)	5.26%
Inflation Rate (hydro)	6.00%	Discount Rate (hydro)	3.77%	Length of Amortization	20

TABLE 12
West Windsor Pollution Control Plant
Capital Cost Comparison - Present Value Analysys
273 000 m3/Day Plant Capacity

		Conventional Activated Sludge		Modified Activated Sludge		Trickling Filter Solids Contact		Rotating Biological Contactor		Biological Aerated Filter	
		Capital	Operating Cost	Capital	Operating Cost	Capital	Operating Cost	Capital	Operating Cost	Capital	Operating Cost
Capital Cost		$67,440,000		$51,020,000		$58,760,000		$59,920,000		$59,110,000	
Annual Payment	Year	$7,921,477	$568,000	$5,992,790	$568,000	$6,901,928	$322,000	$7,038,181	$503,000	$6,943,038	$455,000
Present Value	1	$7,525,403	$580,186	$5,693,151	$580,186	$6,556,831	$328,908	$6,686,272	$513,792	$6,595,886	$464,762
	2	$7,149,133	$592,634	$5,408,493	$592,634	$6,228,990	$335,965	$6,351,958	$524,815	$6,266,092	$474,733
	3	$6,791,676	$605,349	$5,138,068	$605,349	$5,917,540	$343,173	$6,034,360	$536,074	$5,952,788	$484,918
	4	$6,452,093	$618,336	$4,881,165	$618,336	$5,621,663	$350,536	$5,732,642	$547,576	$5,655,148	$495,322
	5	$6,129,488	$631,602	$4,637,107	$631,602	$5,340,580	$358,056	$5,446,010	$559,324	$5,372,391	$505,949
	6	$5,823,014	$645,153	$4,405,251	$645,153	$5,073,551	$365,738	$5,173,710	$571,324	$5,103,771	$516,804
	7	$5,531,863	$658,994	$4,184,989	$658,994	$4,819,873	$373,585	$4,915,024	$583,581	$4,848,583	$527,892
	8	$5,255,270	$673,133	$3,975,739	$673,133	$4,578,880	$381,600	$4,669,273	$596,102	$4,606,154	$539,217
	9	$4,992,506	$687,574	$3,776,952	$687,574	$4,349,936	$389,787	$4,435,809	$608,891	$4,375,846	$550,786
	10	$4,742,881	$702,326	$3,588,105	$702,326	$4,132,439	$398,150	$4,214,019	$621,954	$4,157,054	$562,603
	11	$4,505,737	$717,394	$3,408,700	$717,394	$3,925,817	$406,692	$4,003,318	$635,298	$3,949,201	$574,673
	12	$4,280,450	$732,786	$3,238,265	$732,786	$3,729,526	$415,417	$3,803,152	$648,928	$3,751,741	$587,002
	13	$4,066,428	$748,507	$3,076,351	$748,507	$3,543,050	$424,330	$3,612,994	$662,850	$3,564,154	$599,596
	14	$3,863,106	$764,566	$2,922,534	$764,566	$3,365,897	$433,434	$3,432,345	$677,072	$3,385,946	$612,460
	15	$3,669,951	$780,969	$2,776,407	$780,969	$3,197,603	$442,733	$3,260,727	$691,598	$3,216,649	$625,600
	16	$3,486,453	$797,725	$2,637,587	$797,725	$3,037,722	$452,231	$3,097,691	$706,436	$3,055,816	$639,022
	17	$3,312,131	$814,840	$2,505,707	$814,840	$2,885,836	$461,934	$2,942,806	$721,592	$2,903,026	$652,732
	18	$3,146,524	$832,322	$2,380,422	$832,322	$2,741,544	$471,844	$2,795,666	$737,074	$2,757,874	$666,736
	19	$2,989,198	$850,179	$2,261,401	$850,179	$2,604,467	$481,967	$2,655,883	$752,887	$2,619,981	$681,041
	20	$2,839,738	$868,419	$2,148,331	$868,419	$2,474,244	$492,308	$2,523,089	$769,040	$2,488,982	$695,652
Net Present Value		$96,553,042	$14,302,993	$73,044,725	$14,302,993	$84,125,990	$8,108,387	$85,786,748	$12,666,206	$84,627,081	$11,457,503
			$110,856,035		$87,347,717		$92,234,377		$98,452,955		$96,084,584

Interest Rate	10.00%	Inflation Rate	4.50%	Discount Rate (capital)	5.26%
Inflation Rate (hydro)	6.00%	Discount Rate (hydro)	3.77%	Length of Amortization	20

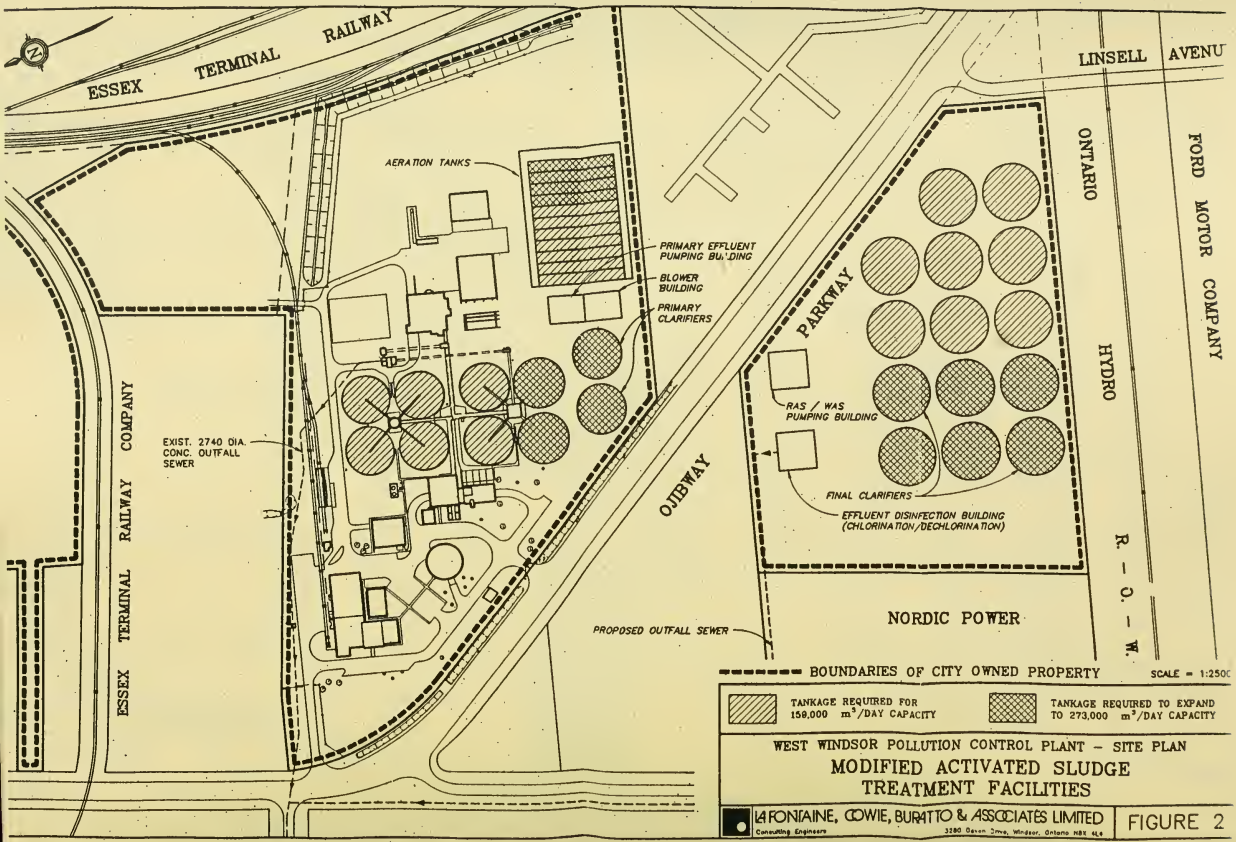

ESSEX TERMINAL RAILWAY
LINSELL AVENU
FORD MOTOR COMPANY
ONTARIO HYDRO R. – O. – W.
AERATION TANKS
PRIMARY EFFLUENT PUMPING BUILDING
BLOWER BUILDING
PRIMARY CLARIFIERS
PARKWAY
OJIBWAY
RAS / WAS PUMPING BUILDING
FINAL CLARIFIERS
EFFLUENT DISINFECTION BUILDING (CHLORINATION/DECHLORINATION)
EXIST. 2740 DIA. CONC. OUTFALL SEWER
ESSEX TERMINAL RAILWAY COMPANY
PROPOSED OUTFALL SEWER
NORDIC POWER
BOUNDARIES OF CITY OWNED PROPERTY
SCALE = 1:250C
TANKAGE REQUIRED FOR 159,000 m³/DAY CAPACITY
TANKAGE REQUIRED TO EXPAND TO 273,000 m³/DAY CAPACITY
N

WEST WINDSOR POLLUTION CONTROL PLANT – SITE PLAN

CONVENTIONAL ACTIVATED SLUDGE
TREATMENT FACILITIES

LAFONTAINE, COWIE, BURATTO & ASSOCIATES LIMITED
Consulting Engineers
3280 Devon Drive, Windsor, Ontario N8X 4L4

FIGURE 1

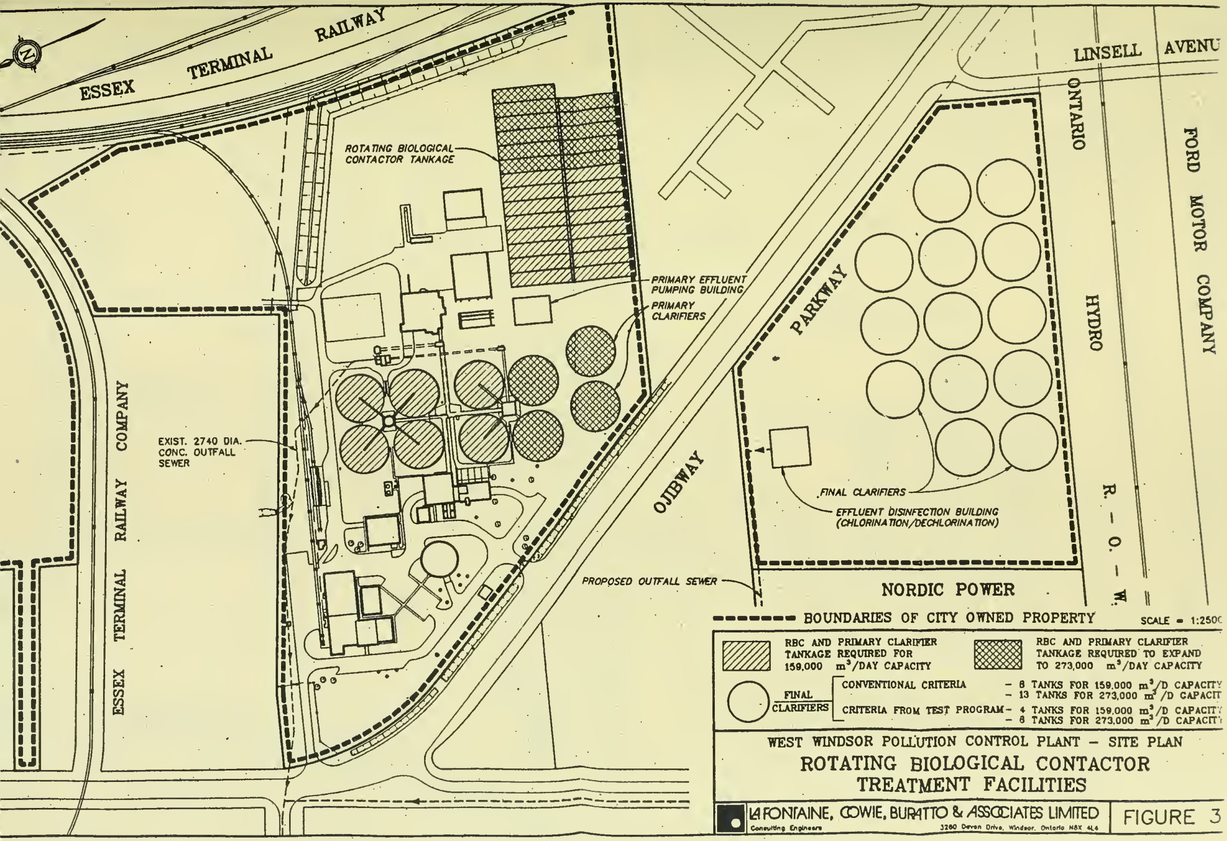

ESSEX TERMINAL RAILWAY
N
LINSELL AVENU
FORD MOTOR COMPANY
ONTARIO HYDRO R. – O. – W.
AERATION TANKS
PRIMARY EFFLUENT PUMPING BUILDING
BLOWER BUILDING
PRIMARY CLARIFIERS
PARKWAY
OJIBWAY
RAS / WAS PUMPING BUILDING
FINAL CLARIFIERS
EFFLUENT DISINFECTION BUILDING (CHLORINATION/DECHLORINATION)
EXIST. 2740 DIA. CONC. OUTFALL SEWER
ESSEX TERMINAL RAILWAY COMPANY
PROPOSED OUTFALL SEWER
NORDIC POWER
BOUNDARIES OF CITY OWNED PROPERTY
SCALE = 1:250
TANKAGE REQUIRED FOR 159,000 m³/DAY CAPACITY
TANKAGE REQUIRED TO EXPAND TO 273,000 m³/DAY CAPACITY

WEST WINDSOR POLLUTION CONTROL PLANT – SITE PLAN

MODIFIED ACTIVATED SLUDGE
TREATMENT FACILITIES

LAFONTAINE, COWIE, BURATTO & ASSOCIATES LIMITED
Consulting Engineers
3280 Devon Drive, Windsor, Ontario N8X 4L4

FIGURE 2

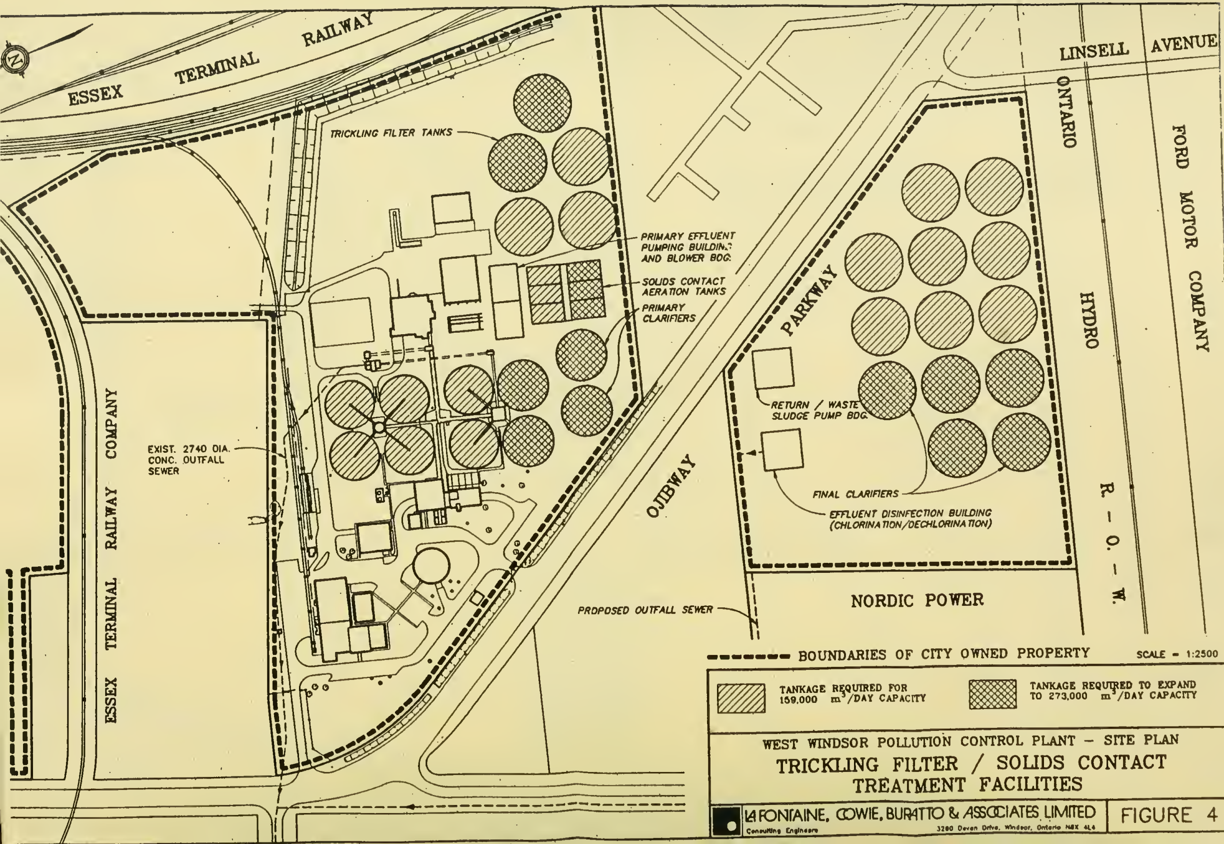

ESSEX TERMINAL RAILWAY
LINSELL AVENU
ONTARIO HYDRO R. – O. – W.
FORD MOTOR COMPANY
ROTATING BIOLOGICAL CONTACTOR TANKAGE
PRIMARY EFFLUENT PUMPING BUILDING
PRIMARY CLARIFIERS
PARKWAY
OJIBWAY
FINAL CLARIFIERS
EFFLUENT DISINFECTION BUILDING (CHLORINATION/DECHLORINATION)
EXIST. 2740 DIA. CONC. OUTFALL SEWER
PROPOSED OUTFALL SEWER
NORDIC POWER
ESSEX TERMINAL RAILWAY COMPANY
BOUNDARIES OF CITY OWNED PROPERTY
SCALE = 1:250C
RBC AND PRIMARY CLARIFIER TANKAGE REQUIRED FOR 159,000 m³/DAY CAPACITY
RBC AND PRIMARY CLARIFIER TANKAGE REQUIRED TO EXPAND TO 273,000 m³/DAY CAPACITY
FINAL CLARIFIERS
CONVENTIONAL CRITERIA – 8 TANKS FOR 159,000 m³/D CAPACITY
– 13 TANKS FOR 273,000 m³/D CAPACIT
CRITERIA FROM TEST PROGRAM – 4 TANKS FOR 159,000 m³/D CAPACITY
– 8 TANKS FOR 273,000 m³/D CAPACITY

WEST WINDSOR POLLUTION CONTROL PLANT – SITE PLAN

ROTATING BIOLOGICAL CONTACTOR TREATMENT FACILITIES

LAFONTAINE, COWIE, BURATTO & ASSOCIATES LIMITED
Consulting Engineers
3280 Devon Drive, Windsor, Ontario N8X 4L4

FIGURE 3


ESSEX TERMINAL RAILWAY
N
TRICKLING FILTER TANKS
PRIMARY EFFLUENT PUMPING BUILDING AND BLOWER BDG.
SOLIDS CONTACT AERATION TANKS
PRIMARY CLARIFIERS
EXIST. 2740 DIA. CONC. OUTFALL SEWER
ESSEX TERMINAL RAILWAY COMPANY
OJIBWAY PARKWAY
PROPOSED OUTFALL SEWER
LINSELL AVENUE
ONTARIO HYDRO R. – O. – W.
FORD MOTOR COMPANY
RETURN / WASTE SLUDGE PUMP BDG.
FINAL CLARIFIERS
EFFLUENT DISINFECTION BUILDING (CHLORINATION/DECHLORINATION)
NORDIC POWER

- - - - BOUNDARIES OF CITY OWNED PROPERTY

SCALE – 1:2500

▨ TANKAGE REQUIRED FOR 159,000 m^3/DAY CAPACITY

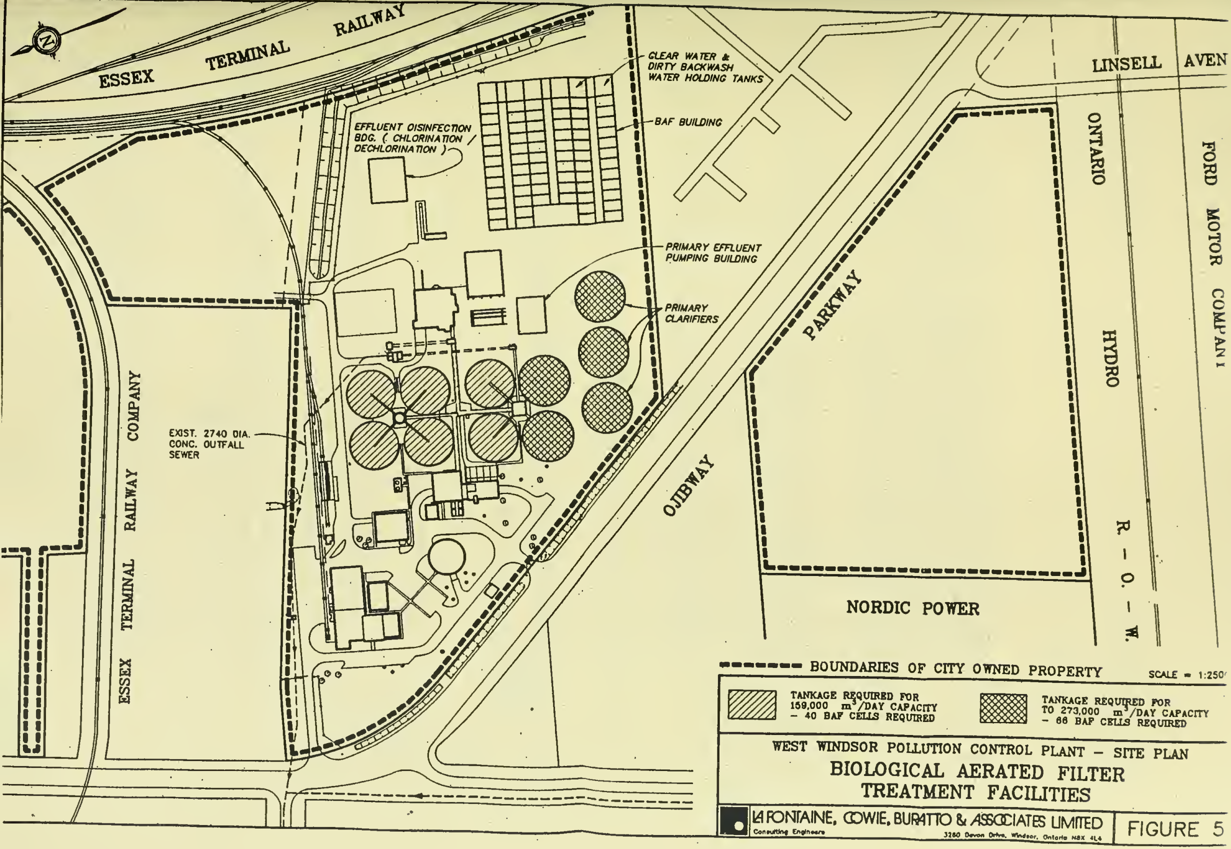
▩ TANKAGE REQUIRED TO EXPAND TO 273,000 m^3/DAY CAPACITY

WEST WINDSOR POLLUTION CONTROL PLANT – SITE PLAN

TRICKLING FILTER / SOLIDS CONTACT TREATMENT FACILITIES

LAFONTAINE, COWIE, BURATTO & ASSOCIATES LIMITED
Consulting Engineers
3280 Devon Drive, Windsor, Ontario N8X 4L4

FIGURE 4



ESSEX TERMINAL RAILWAY
CLEAR WATER & DIRTY BACKWASH WATER HOLDING TANKS
BAF BUILDING
EFFLUENT DISINFECTION BDG. (CHLORINATION / DECHLORINATION)
PRIMARY EFFLUENT PUMPING BUILDING
PRIMARY CLARIFIERS
LINSELL AVEN
ONTARIO HYDRO R. – O. – W.
FORD MOTOR COMPANY
PARKWAY
OJIBWAY
NORDIC POWER
EXIST. 2740 DIA. CONC. OUTFALL SEWER
ESSEX TERMINAL RAILWAY COMPANY
BOUNDARIES OF CITY OWNED PROPERTY
SCALE = 1:250
TANKAGE REQUIRED FOR 159,000 m³/DAY CAPACITY – 40 BAF CELLS REQUIRED
TANKAGE REQUIRED FOR TO 273,000 m³/DAY CAPACITY – 66 BAF CELLS REQUIRED

WEST WINDSOR POLLUTION CONTROL PLANT – SITE PLAN

BIOLOGICAL AERATED FILTER TREATMENT FACILITIES

LA FONTAINE, COWIE, BURATTO & ASSOCIATES LIMITED
Consulting Engineers
3260 Devon Drive, Windsor, Ontario N8X 4L4

FIGURE 5

retention time. Two Humboldt centrifuges 'thicken' digested sludge to ~10 percent TS (higher solids too difficult to handle; some centrifuge plugging was experienced) which is still suitable for land utilization; 60 to 90 day sludge storage is provided. Sludge has been designated (certified) as an organic fertilizer. Digester supernatant and centrate are equalized prior to return to the rock trickling filters.

- Power outages are a problem. The plant load management system is not entirely reliable. A co-generation system is provided and is interconnected to the grid. Power is sold at ~4.25 cents US/kWh. Waste heat is used for digester and building heating.

- When an impending new effluent permit is delivered, effluent bio-monitoring will be required. The plant has had few exceedances in this area; exceedances have generally been due to 'high' NH_3. The receiving stream is a warm water habitat.

- Plant operational staff had design input at the 40. 60 and 80 percent design stages.

6) LITTLETON/ENGLEWOOD REGIONAL WASTEWATER TREATMENT PLANT
ENGLEWOOD, COLORADO

Date:	5 April, 1991; 2:20 p.m.
Contact:	Mr.Dennis Stowe
Attachment:	description of original plant located at this site; plant is currently in re-construction with addition of new TF/SC process.
Plant Design:	37 USmgd; partial seasonal nitrification.

- Feed to the plant is from the two cities named above. A new 20 USmgd pump station was under construction; however the plant design capacity was 37 USmgd. The existing plant bypasses around the existing rock trickling filters on a regular basis.

- Relatively little information was available from this plant because of the reconstruction activity and time available with plant staff. This plant was in some difficulty operationally because of reconstruction activity. One new plastic media TF/SC unit is in construction in parallel with an existing air activated sludge facility. Ultimately, three additional TF/SC units may be built. The TF tower is designed with complete backwashing capability for snail control. Backwash water will be chlorinated to kill snails and then pumped to digestion.

Effluent objectives will be for nitrification with 5.4 mg/L in summer and 13 mg/L winter NH_3-N limits. These values were developed based on a receiving stream study.

The existing activated sludge plant consisted of five parallel trains with GRAY/Sanitaire diffusers. MLSS was high at about 6000 mg/L. Secondary clarifiers showed evidence of a severe incidence of Nocardia type foam/scum with much floating scum on the surface of the clarifiers. An interesting problem observed at the plant was the RAS wet well; it was constructed for 8 clarifiers but only four clarifiers were installed. This resulted in dead zones in the over-sized wet well.

Please contact CH2M HILL for correction of errors or omissions in these notes.

Site Visit Summary prepared by:

CH2M-HILL ENGINEERING LTD.

Joe P. Stephenson, P.Eng.

TENTATIVE SCHEDULE
revised April 1, 1991

WEST WINDSOR PCP UPGRADE

TF/SC SITE VISITS - APRIL 2 TO APRIL 6 ,1991

TENTATIVE SCHEDULE

DAY	EVENT	TIME	DESCRIPTION
TUES. APRIL 2	FLIGHT TO DALLAS	9:00 am DETROIT	DL 1271 FLIGHT DETROIT TO DALLAS DEPARTS 9:00 am ARRIVES 10:45 am
	DRIVE TO GARLAND	11:15 am	DRIVE FROM DALLAS TO GARLAND, 1 h 15 min TRIP (11:15 - 12:30)
	LUNCH	12:30 pm	QUICK LUNCH (12:30 - 1:00)
	PLANT VISIT	1:00 pm	VISIT DUCK CREEK GARLAND WPCP (1:30 - 4:30)
	DINNER	6:00 pm	
	HOTEL	8:00 pm	HOLIDAY INN NORTHEAST ON JUPITER RD, GARLAND, PHONE (214) 341-5400
WED. APRIL 3	BREAK	7:30 am	BREAKFAST FROM 7:30 TO 8:30 am
	PLANT VISIT	9:00 am	VISIT ROWLETT CREEK GARLAND WPCP (9:00 - 12:00)
	LUNCH	12:15 am	LUNCH FROM 12:15 - 1:15 am
	DRIVE TO DALLAS	1:15 pm	DRIVE TO DALLAS AIRPORT (1:15 - 2:30 pm)
	FLIGHT TO SALT LAKE CITY	3:45 pm	DL 1735 FLIGHT FROM DALLAS TO SALT LAKE CITY DEPARTS 3:45 pm, ARRIVES 5:17 pm
	DINNER	7:00 pm	
	HOTEL	9:00 pm	OLYMPUS HOTEL, 161 WEST 600 SOUTH, SALT LAKE CITY, PHONE (801) 521-7373

TENTATIVE SCHEDULE
revised April 1, 1991

DAY	EVENT	TIME	DESCRIPTION
THUR. APRIL 4	BREAK	7:30 am	BREAKFAST FROM 7:30 TO 8:30 am
	PLANT VISIT	9:00 am	VISIT CENTRAL VALLEY WPCP (9:00 - 11:45 pm)
	LUNCH	12:00 pm	LUNCH FROM 12:00 - 1:30 pm
	PLANT VISIT	1:45 pm	VISIT SALT LAKE CITY WPCP (1:45 - 4:30 pm)
	DINNER	6:00 pm	DINNER AT AIRPORT (6:00 - 7:15 pm)
	FLIGHT TO DENVER	7:54 pm	DL 1298 FLIGHT FROM SALT LAKE CITY TO DENVER, DEPART 7:54 pm, ARRIVES AT 9:20 pm
	HOTEL	10:00 pm	LANDMARK BEST-WESTERN HOTEL, 455 SOUTH COLORADO BLVD, DENVER, PHONE (303) 388-5561
FRI. APRIL 5	BREAK	7:00 am	BREAKFAST FROM 7:00 TO 7:45 am
	DRIVE TO BOULDER	8:00 am	DRIVE TO BOULDER WPCP
	PLANT VISIT	9:30 am	VISIT BOULDER WPCP (9:30 - 12:00)
	LUNCH AND DRIVE	12:15 pm	QUICK LUNCH (12:00 - 12:30 pm) AND DRIVE BACK TO DENVER (12:30 - 1:30 pm)
	PLANT VISIT	2:00 pm	VISIT LITTLETON-ENGLEWOOD WPCP (2:00 - 5:00 pm)
	DINNER	7:30 pm	DINNER
	HOTEL	9:00 pm	LANDMARK BEST-WESTERN HOTEL, 455 SOUTH COLORADO BLVD, DENVER, PHONE (303) 388-5561
SAT. APRIL 6	BREAK	7:00 am	BREAKFAST FROM 7:00 - 7:30 am
	FLIGHT TO DETROIT	9:10 am	DL 334 FLIGHT FROM DENVER TO DETROIT (WITH STOPOVER IN CINCINNATI) DEPARTS 9:10 am, ARRIVES CINCINNATI 1:40 pm, DL 3250 DEPARTS 2:20 pm ARRIVES DETROIT AT 3:40 pm

CITY OF WINDSOR EVALUATION OF
INNOVATIVE ALTERNATIVE FOR THE
WEST WINDSOR PCP UPGRADE

SITE VISIT

DATE: May 3, 1991

LOCATION:

PROJECT: ONT29343.C0

ATTENDEES:

Mr. L. Romano, City of Windsor
Mr. H. Horneck, LCBA
Mr. M. Newbigging, CH2M HILL
Mr. M. Maki, City of Windsor
Mr. J. Stephenson, CH2M HILL

SUBJECT: Summary of Site Visit to BAF Facilities in Quebec

A site visit to two BAF (Biocarbone) units located in Quebec was conducted on May 3, 1991. Both plants are designed for organic or BOD_5 removal only. The following memorandum summarizes the main points discussed during the trip.

Sherbrooke, Quebec

The Sherbrooke plant was re-visited since the plant is now starting up. The City and CH2M HILL visited the plant a year ago (May 8, 1990) when the plant was still under construction. Representatives from John Meunier Inc. (Mr. F. Seguin, Mr. H. Van Hoi and Mr. M. Emond) and Le Groupe Teknika (Ms. A. Kantardjieff) showed our group around the plant. The plant had just been started and was going through initial testing. The plant was mainly treating clean water. Our group was toured through the entire facility, the main points discussed were:

- Plant capacity is rated at an average daily flow of 80,000 m^3/d (17.6 MIGD) with a peak capacity of 125,000 m^3/d (27.5 MIGD). The raw wastewater is expected to have a total BOD_5 concentration of approximately 100 mg/L. The plant effluent criteria is 20, 20, and 0.8 for suspended solids, total BOD_5 and total phosphorus.

- Sludge from the plant will be lime stabilized. The plant has a slurry tank into which the plant can treat thickened or co-thickened sludge. Lime can also be used to adjust the pH of the raw sewage.

- Sixty percent of the flow to the plant is by gravity. In addition, two main pumping stations discharge to the plant. The influent works has stainless steel gates, with the ability to control bypass flows.

- The plant has J. Meunier raw screens with 0.5 inch openings. Screens can be controlled by headloss across the screen or on a time elapse basis. The number of screens in service is controlled by the flow rate. Belt conveyors are used for removal of the screened material.

- Aerated grit tanks are used for grit removal. Tanks also has surface skimmers for grease removal. Air flows are regulated along the tanks. High air flow are maintained at head end of the tank, tapering to low air flows at the effluent. Grit tanks are enclosed in a building, all equipment is explosion rated.

- The primary clarifiers are also covered, although not enclosed. The primary clarifiers are used to settled raw sewage and can be used as thickeners (obtained up to 7% TS). Clarifiers are "tube settlers", manufacturer is Degremont. Tanks are provided for chemical and polymer, with mixing tanks for addition prior to primaries. System appears to be very flexible for chemical and polymer addition. In terms of the operation of the thickeners, backwash water can be thickened then proceed to primaries or directly to presses.

- The BAF (Biocarbone) process is enclosed in a building and each train (there are two) of BAF cells is enclosed in a fibreglass encloser. Air is supplied by five blowers, three are centrifugal and two are positive displacement. The centrifugal blowers are Hoffman 300 hp units, with variable frequency controllers and auto suction throttling capability. The centrifugal blowers are used primarily for process air. The positive displacement blowers are used for backwashing. The blowers are provided with insulating curtains for noise reduction. The inlet gates to the BAF cells are controlled. Flow passes over a plate to dissipate the flow.

- During our walk by the BAF cells, the unit was being backwashed. Our group was able to observe the various cycles of air, water and air and water. The operation of the three syphons, located on one wall of the tank, was also observed by the group. It was noted that a significant amount of media was being backwashed out the tank and into the backwash channel. Meunier indicated that this occurs during the initially cleaning cycles, since some of the media fines float on the liquid surface. The large media in the channel is pumped back into the tank after backwashing.

- Our group inspected the backwash pumping gallery. The backwash pump was exhibiting a good deal of vibration, Mr. M. Emond indicated that this problem was being corrected.

- Thickened sludge from the plant is pumped through two plate and frame filter presses. Pumps for sludge feeding and chemical and polymer addition are located in an adjacent gallery.

Boisbriand, Quebec

Mr. F. Seguin, Mr. H. Van Hoi of J. Meunier also took our group on a brief tour of the Boisbriand plant located north of Montreal. The Boisbriand plant is rated at 25,000 m^3/d (5.5) MIGD. The plant is operated partially by J. Meunier, during the tour we met Mr. C. Meunier, who is in charge of the plant operations. The following items were discussed:

- The control room was visited. The computer system provides current status of the BAF unit and trending data. The plant was in the process of conducting an air scouring when we arrived. The air scouring was being used by the operators to breakup the media and move the solids further into the bed. The operators use this procedure to avoid unnecessary backwashing.

- The plant has tube settlers, similar to Sherbrooke. The clarifiers are Degremont equipment and were constructed from the existing "regular" primaries during the last expansion.

- The BAF (biocarbone) system used at this plant is different from Sherbrooke. The cells (eight) are pie shaped and form a circle. Influent flows in on the outside of the circle, effluent flows into the centre of the circle. The water backwashing is performed by a mechanism in the centre well which seals the effluent standpipe and pumps the effluent through the cell being backwashed. The suppliers indicated that they use this setup for smaller plant, and are currently planning a similar facility for Lavel, Quebec. The proposed Lavel plant will have two circular BAF units, each approximately a third larger than Boisbriand's cells.

- The plant was also using an influent online TOD meter, which J. Meunier distributes, to control the process air being used at the plant, in an attempt to minimize air usage and excessive backwashing.